HYDRODYNAMIC STUDY OF GAS-SOLID FLUIDIZATION OF A TERNARY MIXTURE IN A TAPERED BED

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By

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CERTIFICATE

This is to certify that the thesis entitled **Hydrodynamic study of Gas-Solid Fluidization of a ternary mixture in a tapered bed** submitted by **Smruti Ranjan Panda** to National Institute of Technology, Rourkela is a record of bonafide project work under my supervision and is worthy for the partial fulfillment of the degree of Bachelor of Technology (Chemical Engineering) of the Institute. The candidate has fulfilled all prescribed requirements and the thesis, which is based on candidate's own work, has not been submitted elsewhere.

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ABSTRACT

Hydrodynamic characteristics of fluidization in conical or tapered beds differ from those in columnar beds due to the variation of superficial velocity in the axial direction of the beds. In the former, fixed and fluidized regions could coexist and the sharp peaking of the pressure drop could occur, thereby giving rise to a remarkable pressure drop-flow rate hysteresis loop at incipient fluidization. To explore these unique properties, a series of experiments was carried out of homogenous well-mixed ternary mixtures of three different particle sizes at varying compositions in gas-solid tapered fluidized beds with various tapering angles. The tapering angle of the beds has been found to dramatically affect the beds' behavior. Other hydrodynamic characteristics determined experimentally included the maximum pressure drop, minimum velocity of fluidization, bed fluctuation ratio and bed expansion ratio. The dependence of these quantities on average particle diameter, mass fraction of the fines in the mixture, bed height and cone angle has been discussed.

Based on dimensional analysis, correlations have been developed with the system parameters viz., geometry of tapered bed, particle diameter, static bed height, density of solid and gas and superficial velocity of the fluidizing medium. Correlations have been also been developed for bed fluctuation ratio and bed expansion ratio in gas-solid tapered fluidized bed of ternary mixture of irregular particles using factorial design method. Experimental values of critical fluidization velocities and maximum bed pressure drops have been compared with the developed correlations in literature.

Keywords: Gas–solid fluidization, homogeneous ternary mixtures, irregular particles, maximum pressure drop, minimum fluidization velocity, bed fluctuation ratio and bed expansion ratio.

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NOMENCLATURE

D_0 : Bottom diameter of tapered bed, m.

D_1 : Top diameter of tapered bed, m.

D_c : mean diameter of cone, m

dp : particle diameter, m.

h_s : Initial static height of the bed, m.

U : superficial velocity of the fluidizing fluid, m/s.

U_{mf} : Minimum fluidization velocity based on bottom diameter of the bed, m/sec.

G_f : flow rate of fluid at fluidization condition, m^3/hr

G_{mf} : flow rate of fluid at minimum fluidization condition, , m^3/hr

r : bed fluctuation ratio, dimensionless

R : bed expansion ratio, dimensionless

g : gravity ,$ms\text{-}^2$

dp_{sm}: Avg. particle diameter of ternary mixture, m

ΔPmf : Pressure drop through the particle bed at minimum fluidization condition (N/m^2)

Ar : Archimedes number $=\frac{g\ dp_{sm}^3 \rho_{sm}^2}{\mu_f^2}$

Re_c : Critical Reynolds number

X_i : mass fraction of i^{th} component

Fr : froud's number

Greek letters

ρ_f : Fluid density, kg/m^3

α :tapered angle, deg

μ_f : Fluid viscosity, kg/ ms

ρ_{sm} : Density of binary mixture, kg /m^3

ρ_{si} : Solid density, kg/m^3

ε: bed voidage , dimensionless

Φs : average sphericty

Subscripts

mf : minimum fluidzation

f : fluid

c : critical

sm : sauter mean

s : static or stagnant

CHAPTER 1

INTRODUCTION

1. INTRODUCTION

Fluidization is an established fluid-solid contacting technique, which finds extensive applications in combustion, gasification, carbonization, drying of solids, coating of particles, and many others. The fluidized bed reactor is extremely acclaimed in the process industry because of some of its specific features viz. high rate of heat and mass transfer, continuity in operation, no moving parts involved and rapid solid mixing leading to an early isothermal condition throughout the bed.

With the development of fluidized bed coal combustion and the recent interest in the use of fluidized beds for waste utilization and for dry solids separation, the potential applications of multi-component fluidized beds are on the rise. It is because, the fluidized particles though uniform in size at beginning , may change due to the attrition, coalescence and chemical reaction ,thereby affecting the quality of fluidization by high elutriation loss, de-fluidization , segregation in size, inhomogeneous residence time in the bed , leading to non-uniform products for wide particle size distribution. Therefore proper characterization of the bed dynamics for the binary and the multi-component mixtures in gas solid systems is an important prerequisite for their effective utilization, where the combination of particle size ,density and shape influence fluidization behavior [1].

In spite of the various advantages, the efficiency and quality of fluidization is adversely affected .In cylindrical beds, the particle size reduction results in entrainment, limitation of operating velocity in addition to other demerits like slugging, non-uniform fluidization associated with such beds. Various techniques including introducing of baffles, operation in multistage units, imparting vibrations and alteration in bed geometry have been advocated from time to time to tackle the problems. These disadvantages can be overcome by the use of tapered fluidized beds in which superficial velocity of the fluid gradually reduces with height due to increase in cross-sectional area. Tapered fluidized beds have found wide applicability in many industrial processes such as biological treatment of waste water, immobilized bio-film reaction, incineration of waste materials, coating of nuclear fuel particles, crystallization, coal gasification and liquefaction, roasting sulfide ores, and food processing. These beds are also useful for fluidization of materials with wide particle size distribution and for exothermic reactions. Due to the angled walls, random and unrestricted particle movement occurs in a tapered bed with reduced back mixing.

Conical fluidized bed is very much useful for the fluidization of wide distribution of particles, since the cross sectional area is enlarged along the bed height from the bottom to the top, therefore the velocity of the fluidizing medium is relatively high at the bottom, ensuring fluidization of the large particles and relatively low at the top, preventing entrainment of the small particles. Since the velocity of fluidizing medium at the bottom is fairly high, this gives rise to low particle concentration, thus resulting in low reaction rate and reduced rate of heat release. Therefore the generation of high temperature zone near the distributor can be prevented. Due to the existence of a gas velocity gradient along the height of a conical bed, it has some favorable special hydrodynamics characteristics. The conical bed has been widely applied in many industrial processes such as,

1. Biological treatment of waste water,

2. Immobilized bio-film reaction,

3. Incineration of waste-materials,

4. Coating of nuclear fuel particles,

5. Crystallization, roasting of sulfide ores,

6. Coal gasification and liquefaction,

7. Catalytic polymerization,

8. Fluidized contactor for sawdust and mixtures of wood residues and

9. Fluidization of cohesive powder.

CHAPTER 2

LITERATURE REVIEW

2.1. THE PHENOMENON OF FLUIDIZATION

On passing fluid (gas or solid) upward through a bed of fine particles, at a low flow rate fluid merely percolates through the void spaces between stationary particles. This is a fixed bed. With an increase in flow rate, particles move apart slightly and a few are seen to vibrate and move about in restricted regions. This is the expanded bed. At a still higher velocity, the pressure drop through the bed increases. At a certain velocity, the pressure drop through the bed reaches the maximum and a point is reached when the particles are all just suspended in the upward flowing gas or liquid. At this moment, the particles at the bottom of the bed begin to fluidize, thereafter the condition of fluidization will extend from the bottom to the top and the pressure drop will decline fairly sharply. Evidently, fluidization is initiated when the force exerted between a particle and fluidizing medium counterbalances the effective weight of the particle, the vertical component of the compressive force between the adjacent particles disappears, and the pressure drop through any section of the bed about equals the weight of fluid and particles in that section. The bed is considered to be just fluidized and referred to as an incipiently fluidized bed or a bed at minimum fluidization. There is a drop in pressure drop after minimum fluidization condition due to restructuring of the bed with increase in bed voidage at lower levels of the bed. Under the assumption that friction is negligible between the particles and the bed walls, also it is assumed that the lateral velocity of fluid is relatively small and can be neglected and the vertical velocity of the fluid is uniformly distributed on the cross sectional area.

Gas–solid systems generally behave in quite different manner. With an increase in flow rate beyond minimum fluidization, large instabilities with bubbling and channeling of gas are observed. At higher flow rates agitation becomes more violent and the movement of solids becomes vigorous. In addition, the bed does not expand much beyond its volume at minimum fluidization. Such a bed is called an aggregative fluidized bed, a heterogeneous fluidized bed, a bubbling fluidized bed, or simply a gas fluidized bed.

2.2. EXPERIMENTAL PHENOMENA

FLOW REGIMES

A typical diagram of the hydrodynamic characteristics of the conical bed is shown in Fig.1 with the increase of superficial gas velocity, U_{g0}, the total pressure drop, ΔP_t, varies along the line of O→A→B→C as given by S. Jing et al [2]. In the different stages, the hydrodynamic characteristics of fluidization of the conical bed are as follows:

O→A stage: Because U_{go} is relatively low, the stagnant height of the particle bed remains unchanged as at the beginning. The total pressure drop, ΔPt, increases up to the maximum point, ΔP_{max}, i.e. point A .This phenomenon is the same as that observed for liquid–solid tapered beds by Peng and Fan [3], and the flow regime is also termed the fixed bed regime. The superficial gas velocity, to which point A in Fig.1 corresponds, is called the minimum fluidized velocity, U_{mf}.

A→.B stage: When U_{go} is higher than U_{mf} , ΔP_t decreases with the increase of U_{go}, and it is observed that the stagnant height of the conical bed does not change. The same phenomenon is also observed for gas–solid systems by Olazar et al. [5] and for liquid–solid systems by Peng and Fan [3]. Here, the flow regime is named a partially fluidized bed. When U_{go} reaches U_{ms}, the characteristics of total pressure drop are different from those in the above two stages.

B→C stage: If U_{go} is greater than U_{ms}, ΔP_t stays nearly constant as shown in below Fig. In this stage, it is observed that slugging fluidization, bubble fluidization and spouting fluidization occur for the conical bed . In this stage, the characteristics of fluidization of the gas–solid conical bed are different from that of liquid–solid ones as reported by Peng and Fan [3]. Depending on the cone angle, the flow regime is called a slugging or spouting fluidization regime.

Reversing the fluidization process, the fluidized bed is de-fluidized by decreasing the superficial gas velocity. The same regimes are observed in below Fig1.

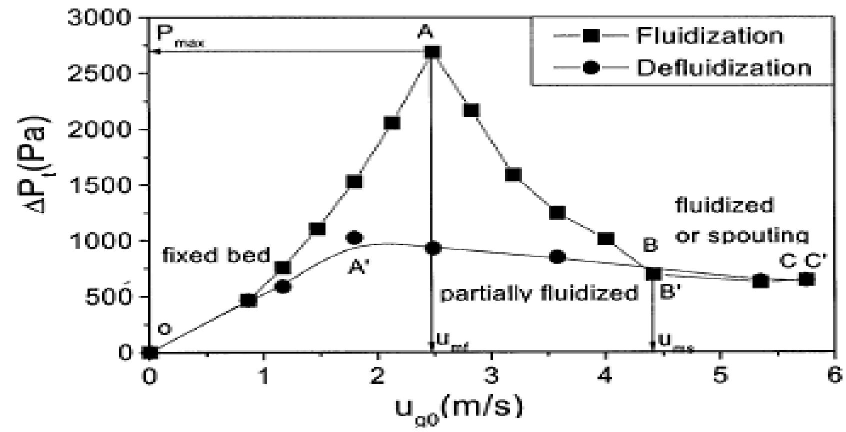

Figure 2.1 Effect of gas velocity on total pressure drop

2.3. DEVELOPMENT OF THE MODEL

In the course of experiments it has been observed that, at a particular velocity, the pressure drop reaches a maximum and the particles in the bed are lifted slightly upward by the fluid. This is followed by the particles at the bottom of the bed beginning to fluidize. Once the particles are unlocked there is a sharp decline in the pressure drop. Evidently, fluidization is initiated when the force exerted by the fluidizing medium flowing through the bed is equal to the total effective weight of the particles in the bed. It is assumed that the lateral velocity of the fluid is relatively small and can be neglected. The pressure drop through a packed bed over a differential height of dh is given by Ergun [6] as follows:

$$\frac{\Delta P}{H} = \frac{150\, U_{mf}\, \mu_g (1-\varepsilon_{mf})^2}{\Phi_s^2\, dp_{sm}^2 \varepsilon_{mf}^3} + \frac{1.75\, U_{mf}^2 \rho_g (1-\varepsilon_{mf})}{\Phi_s dp_{sm} \varepsilon_{mf}^3} \qquad (2.1)$$

i.e. $dP = (AU + BU^2)\, dh$

Where, $A = \frac{150(1-\varepsilon)^2\, \mu_g}{dp_{sm}^2 \varepsilon^3 \Phi_s^2}$ and $B = \frac{1.75\, \rho_f (1-\varepsilon)}{dp_{sm} \varepsilon^3 \Phi_s}$

The overall pressure drop across the bed height, H, is obtained by integrating equation (2.1),

$$-(\Delta P) = \int_{ho}^{H+ho} -(dp) = \int_{ho}^{H+ho} (AU + BU^{\wedge}2)\, dh \qquad (2.2)$$

Now for a conical bed with apex angle of α, we have

$\frac{U}{Uo} = \frac{D_o^2}{D^2}$, $\frac{D}{Do} = \frac{h}{h_o}$ and $U = Uo\frac{h_o^2}{h^2}$

So equation (2) becomes

$$-(\Delta P) = \int_{ho}^{H+ho} (A.\, Uo\frac{c}{h^2} + B.\, U_o^2 \frac{h_o^4}{4})\, dh$$

$$= A.\, U_O . \frac{H.ho}{(H+ho)} + \frac{BU_o^2 . h_o}{3} \left[\frac{(H+h_o)^3}{(H+h_o)^3} - \frac{h_o^3}{(H+h_o)^3}\right] \qquad (2.3)$$

Where $h_o = \frac{D_o}{2\tan\alpha/2}$

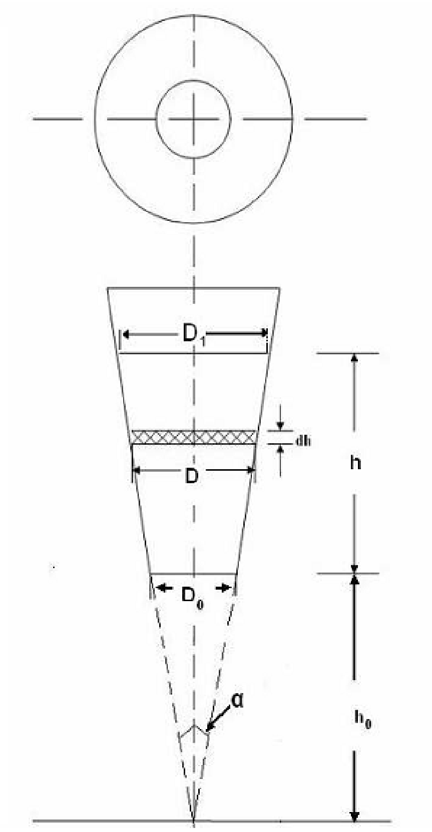

Figure 2.2 Structure of the bed

The area of cross-section-of a conical bed increases continuously from the bottom to the top. So the force exerted by the fluidizing medium on the solid particles is not directly proportional to the pressure drop. The force in a differential bed height of dh is equal to the product of the pressure drop through it, - (dp), and the cross sectional area.

On integration for a conical bed, we get

$$G = \frac{K[(H+h_0)^3 - h_0^3]}{3} = C' \text{ (say)}$$

Where $K = \frac{g(1-\varepsilon)(\rho_s - \rho_f)\pi D_o^2}{4h_o^2}$

Now according to the proposed model as given by Agarwal et al [7] , the particles at the bottom start fluidizing when F=G. therefore the minimum fluidization velocity, U_{mf} can be found out by equating F =G, that gives

$$A_1 U_{mf}^2 + B_1 U_{mf} = C_1$$

$$\text{Or,} U_{mf} = \frac{-B1 + \sqrt{B1^2 + 4 A_1 C_1}}{2A_1} \qquad (2.4)$$

$$\text{Where, } A_1 = \frac{\pi B D_{0H}^3}{4(D_0 + 2H \tan \alpha/2)}, \ B_1 = \frac{\pi A D_0^2 H}{4} \text{ and } C_1 = \frac{K[(H+h_0)^3 - h_0^3]}{3} \qquad (2.5)$$

Putting in equation (3) $U_o = U_{mf}$

$$-(\triangle P_{mf}) = AHh_o U_{mf} + \frac{Bh_0}{3}[(H+h_0)^3 - h_0^3]\frac{U_{mf}}{(H+h_0)^3} \qquad (2.6)$$

2.4. DEVELOPED CORRELATIONS FOR U_{mf} AND ΔP_{mf}

Most of the gas–solid fluidization behavior studies that have been reported are for straight cylindrical or columnar fluidized beds, although a considerable proportion of the fluidized beds have inclined walls or have a tapered bottom section. A velocity gradient exists in the axial direction, leading to unique hydrodynamic characteristics. In spite of its advantages and usefulness, not much work has been reported in literature for understanding certain important characteristics, especially minimum fluidization velocity and maximum pressure drop. Studies have been reported by researchers to determine the factors affecting minimum fluidization velocity and maximum pressure drop. But some of these results are limited to regular particles only. Some of the previous investigations include fixed bed pressure drop calculations, flow regimes, incipient condition of fluidization, voidage distribution and bed expansion calculations and development of a model for maximum pressure drop at incipient fluidization condition of a tapered fluidized bed.

The model developed by Shi et al is based on Ergun's equation [6] and neglects friction between the particles and the wall. Biswal et al [8][9][10] developed theoretical models, for minimum fluidization velocity and pressure drop in a packed bed of spherical particles for gas–solid systems in conical vessels. Due to the angled walls, random and unrestricted particle movement occurs in a tapered bed with reduced back mixing. Olazer et al [5] compared their experimental results with that calculated using the models developed by Gelperin et al. and Gorshtein and Mukhlenov for maximum pressure drop and found that the predictions were not very accurate. They therefore proposed a modified equation for calculation of maximum pressure drop. Later, Peng and Fan [4] made an in-depth study of the hydrodynamic characteristics of solid–liquid fluidization in a tapered bed and derived theoretical models for the prediction of minimum fluidization velocity and maximum pressure drop, based on the dynamic balance of forces exerted on the particle. The experiments were however carried out for spherical particles only. Jing et al. and Shan et al.[2][3] proposed models for gas–solid conical fluidized beds for spherical coarse and fine particles based on the Peng and Fan models but neglected the pressure drop due to the kinetic change in the bed.

Biswal et al. (1982, 1984, and 1985) [8][9][10] developed correlations for fluctuation ratio in a packed bed for spherical and non-spherical particles for gas-solid system in conical vessels but have not included the effect of density of solid and the fluidizing medium.

Some of the well known correlations available for predicting the minimum fluidization velocity (U_{mf}) and maximum pressure drop (P_{max}) for tapered beds are those by Peng and Fan and Jing et al. Peng and Fan developed a model for estimating minimum fluidization velocity and maximum pressure drop for solid–liquid system and spherical particles based on the dynamic balance of forces exerted on the particle. The correlation reported by them for minimum fluidization velocity is given in Eq. (2.7).

$$\mathrm{C_1 U_{mf}} + \mathrm{C_2}\left(\frac{\mathrm{D_0}}{\mathrm{D_1}}\right)\mathrm{U_{mf}^2} - (1-\varepsilon_0)\left(\rho_s - \rho_f\right)g\,\frac{\mathrm{D_0^2 + D_0 D_1 + D_1^2}}{\mathrm{3D_0^2}} = 0 \qquad (2.7)$$

The equation for pressure drop has been developed from Ergun’s equation, which also includes the pressure drop due to a kinetic energy change in the bed:

$$-\triangle P_{max} = C_1 H_s \frac{D_O}{D_1} U_O + C_2 H_s D_O \frac{(D_O^2 + D_O D_1 + D_1^2) U_O^2}{3D_O^3} + (1/2)\left(\frac{U_O}{\varepsilon_O}\right)^2\left[\left(\frac{D_O}{D_1}\right)^4 - 1\right] \qquad (2.8)$$

Jing et al. developed a model based on Ergun's equation for pressure drop calculation but neglecting the pressure drop due to the kinetic energy change in the bed. The equation proposed by them for nearly spherical particles is

$$- \triangle P_{max} = C_1 H_s \frac{D_O}{D_1} U_O + C_2 H_s D_O \frac{(D_O^3 + D_O D_1 + D_1^3) U_O^2}{3 D_1^3} \qquad (2.9)$$

Recently, Sau et.al. [11]developed a correlation for maximum pressure drop in gas–solid tapered fluidized beds

$$Fr = 0.2714(Ar)^{0.3197}(\sin\alpha)^{0.6092}(\frac{\varepsilon_O}{\Phi_s})^{-0.6108} \qquad (2.10)$$

$$\triangle P_{max} = 7.457(\frac{D_1}{D_0})^{0.038}(\frac{dp}{D_0})^{0.222}(\frac{H_s}{D_0})^{0.642}(\frac{\rho_s}{\rho_f})^{0.723} \qquad (2.11)$$

On the basis of Ergun's equation and Baskakov and Gelperm's modification for cone geometry, a packed bed pressure drop equation for conical beds was developed by Biswal.et.al.[10], for gas-solid systems

$$- \triangle P_{mf} = \cos\left(\frac{\alpha}{2}\right)\left[37.17(\tan\alpha)^{-0.47}\frac{\mu_g(1-\varepsilon)^2 R_O(R-R_O)}{dp_{sm}^2 \varepsilon R} + \frac{0.75\rho_f(1-\varepsilon)R_O[1-(R_O/R)^3]U_O^2}{3 dp_{sm}\varepsilon^3}\right] \qquad (2.12)$$

In order to develop correlations for ternary systems, it is necessary to define the particle diameter and the density of the ternary systems. In this study, they are defined as in Goossens et al. [12].

$$Dp_{sm} = \frac{1}{\sum_i \frac{X_i}{dp_i}} \qquad (2.13)$$

$$\frac{1}{dp_{sm}\rho_{sm}} = \frac{w_1}{dp_1 \rho s_1} + \frac{w_1}{dp_1 \rho s_2} + \frac{w_1}{dp_1 \rho s_3} \quad \text{In case of heterogeneous ternary mixture} \qquad (2.14)$$

$$\frac{1}{dp_{sm}} = \frac{w_1}{dp_1} + \frac{w_2}{dp_2} + \frac{w_3}{dp_3}$$ In case of homogeneous ternary mixture (2.15)

The dimensionless correlation for critical fluidization velocity of mixture of irregular particles and maximum pressure drop are given by Sau et al [13],

$$Re_c = 301.416(Ar)^{0.1272}\left(\frac{\rho_s - \rho_f}{\rho_f}\right)^{0.3507}\left(\frac{dp_{sm}}{D_O}\right)^{1.562}(\sin\alpha)^{0.1917} \quad (2.16)$$

$$-\frac{(\triangle P_{\max})}{\rho_{sm} g H_s} = 0.0204(Ar)^{0.20}\left(\frac{H_s}{D_O}\right)^{-0.2273}(\sin\alpha)^{0.2947} \quad (2.17)$$

$$Fr = \left(\frac{U_{mf}}{\sqrt{g dp_{sm}}}\right) \quad (2.18)$$

The average sphericity for the ternary mixture has been calculated from the correlation of Narsimhan [14] for binary and ternary mixture. The equation can be written as

$$\frac{1-\varepsilon}{\Phi_s} = 0.231 \log dp_{sm} + 1.417 \quad (2.19)$$

where dp is the average particle diameter in feet.

The original correlations of Bilbao et al. and Chiba et al.[15] as reported by Clarke et al.[16] for completely mixed bed of homogeneous binary mixture of particles are given by,

$$U_{mf}^{m} = U_{mf}^{c} - \left(U_{mf}^{c} - U_{f}^{f}\right)X_f \quad (2.20)$$

Where Xf, the real volume fraction of fines in the binary mixture is same as the mass fraction of fines in the mixture for different particles of same density: For mixture of particles of same density d is simply the mass mean particle size dp, mm. Eq. (20) for ternary mixture of particles can be written as

$$U_{mf}^{m} = U_{mf}^{c} - \left(U_{mf}^{c} - U_{f}^{i}\right)X_i - (U_{mf}^{c} - U_{f}^{f})X_f \tag{2.21}$$

The critical fluidization velocity is given by Sau et al [17] for binary mixture of regular particles as,

$$Re_c = 16.364(Ar)^{0.1969}\left(\frac{\rho_s-\rho_f}{\rho_f}\right)^{0.3507}\left(\frac{dp_{sm}}{D_O}\right)^{1.1854}(\sin\alpha)^{0.0585} \tag{2.22}$$

$$-\frac{(\Delta P_{\max})}{\rho_{sm} g H_s} = 0.081(Ar)^{0.1493}\left(\frac{H_s}{D_O}\right)^{-0.2337}(\sin\alpha)^{0.3826} \tag{2.23}$$

A mathematical model for the minimum fluidization velocity and maximum bed pressure drop was developed Agarwal and Roy et al [7] given by,

$$U_{MF} = \frac{-B1 + \sqrt{B1^2 + 4\,A_1 C_1}}{2A_1} \tag{2.24}$$

$$A_1 = \frac{\pi B\, D_{0H}^3}{4(D_0+2H\ \tan\alpha/2)} \quad , B_1 = \frac{\pi A D_0^2 H}{4}\, and\ \ C_1 = \frac{K[(H+h_0)^3 - h_0^3]}{3}$$

Where $A = \frac{150(1-\varepsilon)^2\,\mu_g}{dp_{sm}^2\varepsilon^3\Phi_s^2}$, $B = \frac{1.75\,\rho_f(1-\varepsilon)}{dp_{sm}\varepsilon^3\Phi_s^3}$ $and\ K = \frac{g(1-\varepsilon)(\rho_s-\rho_f)\pi D_o^2}{4h_o^2}$

The pressure drop is given as

$$-(\Delta\,\mathrm{P_{mf}}) = \mathrm{AHh_oU_{mf}} + \frac{\mathrm{Bh_0}}{3}[(\mathrm{H}+\mathrm{h_0})^3 - \mathrm{h_0^3}]\frac{\mathrm{U_{mf}}}{(\mathrm{H}+\mathrm{h_0})^3} \tag{2.25}$$

2.5. DEVELOPED CORRELATIONS FOR BED FLUCTUATION RATIO(r)

For gas flow more than the minimum fluidization velocity, the top of the fluidized bed may fluctuate considerably. The extent of the fluctuation and its estimation becomes important while specifying the height of a fluidizer. The fluctuation may be defined as the ratio of the highest and the lowest level of the top of the bed for any fluidizing gas mass velocity. This ratio is termed as the fluctuation ratio. Bed fluctuation and fluidization quality being inter-related, consistent efforts have been made to correlate fluctuation ratio in terms of static and dynamic parameters of the system.

Experimentally it is given by,

$$r = \frac{highest\ level\ the\ top\ of\ bed\ occupies(h_2)}{lowest\ level\ the\ top\ of\ bed\ occupies(h_1)}$$

A correlation for fluctuation ratio in conical vessels for regular particle has been developed by Biswal et al[18] using dimensional analysis approach based on four dimensionless groups neglecting the effect of density of gas and solid particles. The correlation reported by Biswal et al (1984) for fluctuation ratio of regular particle is given in equation (2.26) and for irregular particle is in equation (2.27).

$$r = 3.168[\left(\frac{dp}{D_O}\right)^{0.14}\left(\frac{d_c}{h_s}\right)^{-0.16}\left(\frac{h_S}{D_o}\right)^{-0.24}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.17}] \tag{2.26}$$

The bed fluctuation ratio for irregular particles in conical vessels is given by Biswal et al [19] as

$$r = 9.48[\left(\frac{dp}{D_O}\right)^{0.27}\left(\frac{d_c}{h_s}\right)^{-0.83}\left(\frac{\rho_S}{\rho_f}\right)^{-0.15}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.32}] \tag{2.27}$$

Another set of correlations given by Singh and Roy et al [20] for bed fluctuation ratio in conical conduits are,

For heterogeneous and spherical particles:-

$$r = 0.44 \left(\frac{G_f}{G_{mf}}\right)^{0.58} \left(\frac{\rho_{sm}}{\rho_f}\right)^{-0.10} \left(\frac{D_P}{D_O}\right)^{-0.06} (\tan \alpha)^{-0.17} \tag{2.28}$$

For non-spherical particles:-

$$r = 3.42 \left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{-0.02} \left(\frac{D_o}{dp_{sm}}\right)^{-0.180} \left(\frac{h_S}{D_O}\right)^{0.033} \left(\frac{\rho_{sm}}{\rho_f}\right)^{-0.023} \tag{2.29}$$

For homogeneous and spherical particles:-

$$r = 9.8 * 10^{-2} \left(\frac{G_f}{G_{mf}}\right)^{1.068} \left(\frac{D_o}{dp_{sm}}\right)^{-1.97} \left(\frac{h_S}{D_O}\right)^{-0.25} (\tan \alpha)^{-0.25} \tag{2.30}$$

For any bed the general form for the bed fluctuation ratio is given by Singh et al [20][21],

$$r = K(\frac{dp}{d_c})^a [(\frac{d_c}{h_s})(\frac{\rho_f}{\rho_s})]^b (\frac{G_f - G_{mf}}{G_{mf}})^c \tag{2.31}$$

Table 2.1: values of k, a, b, c for various types of bed for bed fluctuation ratio

Type of bed	K	a	b	c
Cylindrical	1.92	0.04	0.04	0.05
Semi-cylindrical	2.32	0.05	0.04	0.07
Hexagonal	2.3	0.06	0.05	0.06
Square	2.5	0.09	0.04	

2.6. DEVELOPED CORRELATIONS FOR BED EXPANSION RATIO (R)

Expansion of gas-solid fluidized beds may in general result from the volume occupied by bubbles and from increase in voidage of the dense phase. It is given by the expression,

$$R = \frac{h_1 + h_2}{h_s} = \frac{h_{avg}}{h_s}$$

For any bed the general form for the bed expansion ratio is given by Singh et al [20][21] as

$$R = K(\frac{dp}{d_c})^a(\frac{d_c}{h_s})^b(\frac{G_f - G_{mf}}{G_{mf}})^c \qquad (2.32)$$

Table 2.2: values of k, a, b, c for various types of bed for bed expansion ratio

Type of beds	K	a	b	c
Non-spherical :-				
cylindrical	2.55	0.11	0.727	0.433
Semi-cylindrical	5.46	0.26	0.03	0.21
Hexagonal	2.422	0.12		0.35
square	6.09	0.24		0.27
Spherical:-				
Semi-cylindrical	2.92	0.14	0.36	0.16
Hexagonal	2.82	0.13		0.22

CHAPTER 3

MATERIALS AND METHODS

3. 1. <u>EXPERIMENTAL DETAILS</u>

The schematic diagram of experimental setup is shown below.

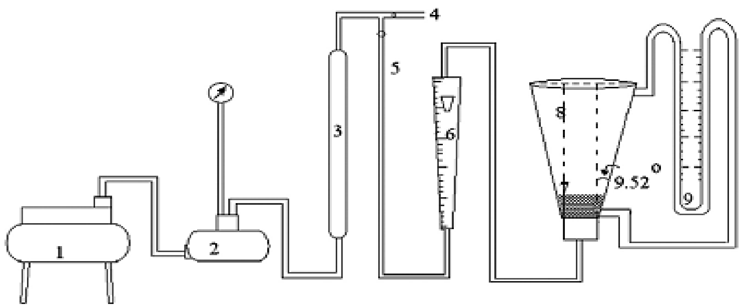

Figure 3.1 Schematic diagram of the experimental set-up 1.Compressor 2. Receiver 3. Silica gel tower4.By pass valve 5.Line valve 6.Rotameter 7.Bed materials 8. Fluidizer 9. Pressure tapping to manometer

Figure 3.2 The experimental set-up for study of hydrodynamics study gas solid fluidization of a ternary mixture in a tapered bed

3.2. CONSTITUENTS OF EXPERIMENTAL SETUP

The experimental set up consists of the following parts,

Air Compressor

It is multistage air compressor of sufficient capacity.

Air Accumulator

It is horizontal cylinder used for storing the compressed air from compressor. There is one G.I. pipe inlet to the accumulator and one by-pass from one end of the cylinder. The exit line also at G.I line taken from the central port of the cylinder. The purpose for using air accumulator in the line is to dampen the pressure fluctuations. The accumulator is fitted with a pressure gauge; the operating pressure in the cylinder is kept at 20psig.

Silica-Gel Column

A silica gel column is provided in the line immediately after the air accumulator to arrest the moisture carried by air from the accumulator.

Pressure Gauge

A pressure gauge in the required range (1-50psig.) is fitted in the line for measuring the working pressure.

Rotameter

Rotameter is used for the measurement of flow rate of air. Two rotameters, one for the lower range (0-20 m3/hr) and the other for the higher range (20-120 m3/hr) were used to measure the air flow rates.

Air Distributor

A 60 mesh screen at the bottom served as the support as well as the distributor. The distributor is an integral part of calming section where it is followed by a conical section. The inside hollow space of the distributor filled with glass beads of 1.5 cm outer diameter, for uniform air distribution.

Conical Fluidizer

The tapered columns were made of Perspex sheets to allow visual observation with different tapered angles. The inlet diameters were 48 mm, 42 mm , 50 mm and 45 mm, where as the outlet diameters were 132mm, 174 mm, 135 mm and 245 mm respectively. The reactor heights were 520 mm, 504 mm, 470 mm and 510 mm respectively. Two pressure tapings are provided for noting the bed pressure drop.

Quick Opening Valve and Control Valve

A globe valve of 1.25cm inner diameter attached to next to the pressure gauge for sudden release of the line pressure. A gate valve of 15mm inner diameter is provided in the line to control the airflow to the bed.

Manometer Panel Board

One set of manometer is arranged in this panel board to measure the pressure drop. Carbon tetrachloride (density=1584 kg m-3) is used as manometer liquid.

3.3. EXPERIMENTAL PROCEDURE

The experiments were carried out in different columns having tapered angles of 4.61°, 7.47° and 9.52° and 11.2°. Three closely sieved samples of dolomite (density=2800 kg m^{-3}) were used for the investigation. For ternary mixture, fairly good mixing has been achieved by coning and quartering method as done in experimental practice and classification has been avoided since the ratio of the largest to the smallest particle size in the mixture is kept below 2.3. Details of the tapered columns are given in Table 4. The densities of the particles were obtained by dividing the weight of the particles by the displaced water volume, when the particles were put into a cylindrical column filled with water.

The above three particle sizes have been mixed in the ratio of50:35:15, 40:30:30, 30:25:45and 20:20:60. The weighed quantity of each solid material of the mixture was poured into the fluidization column. Prior to recording any data the charge was vigorously fluidized with air at a velocity at which entrainment was not observed. After a certain time, the air flow was suddenly stopped to obtain mixed packed bed and then the experiment was started. The initial static bed height was recorded. Then the velocity of the air was increased incrementally allowing sufficient time to each a steady state. The rotameter and manometer readings were noted for each increment in flow rate from which superficial gas velocities and pressure drops were calculated. The velocity at which the pressure drop was maximum was taken as the critical fluidization velocity. The same process was repeated for different initial static bed heights, different mixture of particles and different tapered angles of the tapered beds.

TABLE 3.1: SCOPE OF THE EXPERIMENT

(A)Properties of bed material				(B)Ternary mixture properties		
materials	Particle size ($dp * 10^3$ m)	ρ_s (kg $/m^3$)	Particle Size ratio	mixture	composition	Avg. particle size($dp * 10^3$)
Dolomite(dp_1)	1.55	2800	$dp_1/dp_2 = 1.29$	Mixture-1	50:35:15	1.29
dolomite(dp_2)	1.20	2800	$dp_2/dp_3 = 1.29$	Mixture-2	40:30:30	1.20
dolomite(dp_3)	0.925	2800	$dp_1/dp_3 = 1.68$	Mixture-3	30:25:45	1.125
				Mixture-4	20:20:60	1.106
(C) Bed Parameter						
$H_s(m)$	0.10	0.125		0.15	0.175	
(D) Tapered angle						
α(in degrees)	4.61	5.13		7.47	11.2	

TABLE 3.2: dimensions of available tapered beds

Dimension	Tapered angle(α),in degrees			
	4.61	5.13	7.47	11.2
Bottom diameter, mm	48	50	42	45
Top diameter, mm	132	135	174	245
Height of column, mm	520	470	504	510

CHAPTER 4

RESULTS AND DISCUSSIONS

4.1. FROM EXPERIMENTAL DATA

(A) **STUDY OF MINIMUM FLUIDIZATION VELOCITY AND PRESSURE DROP AT MINIMUM FLUIDIZATION**

The hydrodynamics characteristics of a tapered bed like U_{mf} & ΔP_{mf} is found to be a bit different from the conventional cylindrical conduits. From the experimental observations given from Table A.1 to Table A.22 (refer appendix), following effects of the controlling variables on the hydrodynamic properties are studied.

(1) EFFECT OF BED HEIGHT

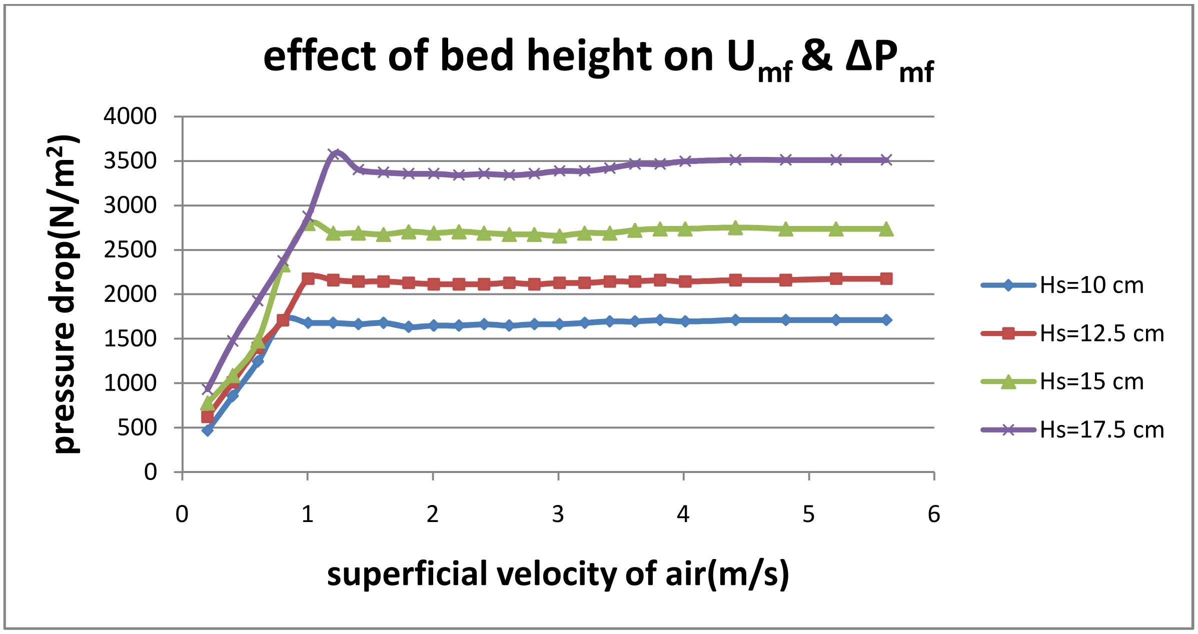

Figure 4.1 Variation of bed pressure drop with gas superficial velocity for cone angle of 7.47° and 40:30:30 mixtures for different Hs.

DISCUSSIONS

It can be seen that there are three regimes namely fixed bed, partially fluidized and fully fluidized bed respectively, and the values of ΔP_{mf} and U_{mf} increase with increasing stagnant bed height. Pressure drop occurs across the bed due to frictional resistance at particle surface and sudden expansion and contraction of flow through interstities among the particles. Also unlike cylindrical conduits, it has a significant peak in minimum fluidization condition where bed changes from quiescent state to homogeneous smoothly expanding condition.

(2) EFFECT OF AVG. PARTICLE DIAMETER

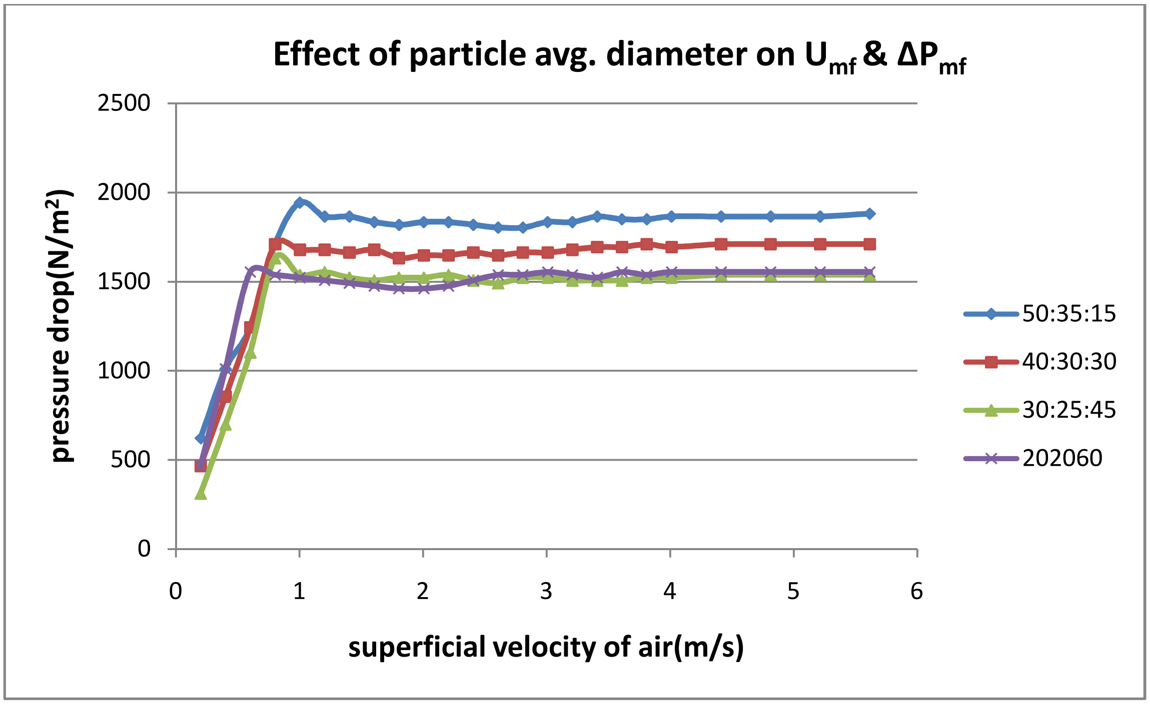

Figure 4.2 Variation of bed pressure drop with gas superficial velocity for cone angle of 7.47° and Hs = 10 cm for different mixtures.

DISCUSSIONS

It is found that with inc. in avg. particle diameter of the ternary mixture, the pressure drop as well as the U_{mf} value increases. In packed bed, different trend is observed from fluidized bed as some fine particles may occupy the interstitial space between the bed mass and the gas finds less space to flow, hence increasing the pressure.

(3) EFFECT OF TAPERED ANGLE

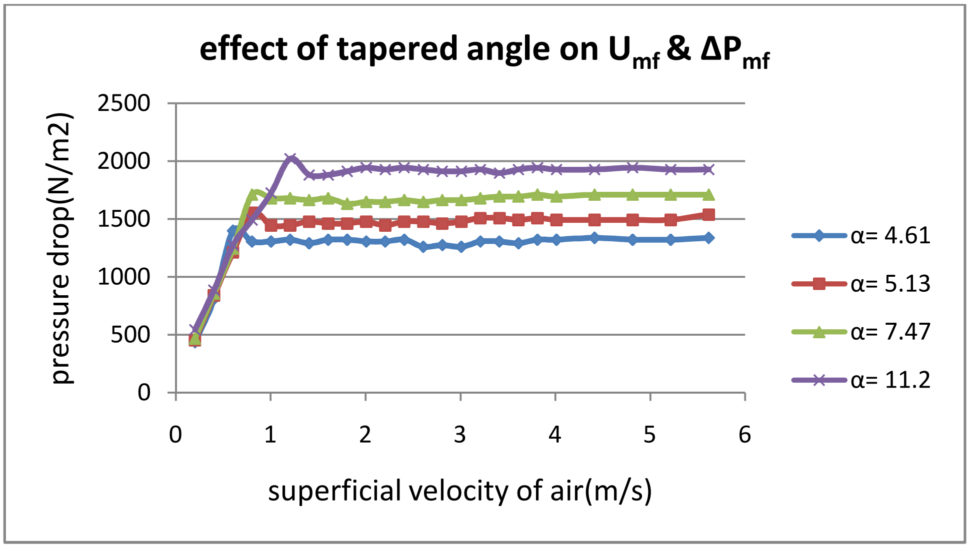

Figure 4.3 Variation of bed pressure drop with gas superficial velocity for varying cone angles of const. H_s = 10 cm and 40:30:30 mixture.

With increase tapered angle, U_{mf} & ΔP_{mf} is found to increase

(B) **STUDY OF BED FLUCTUATION RATIO**

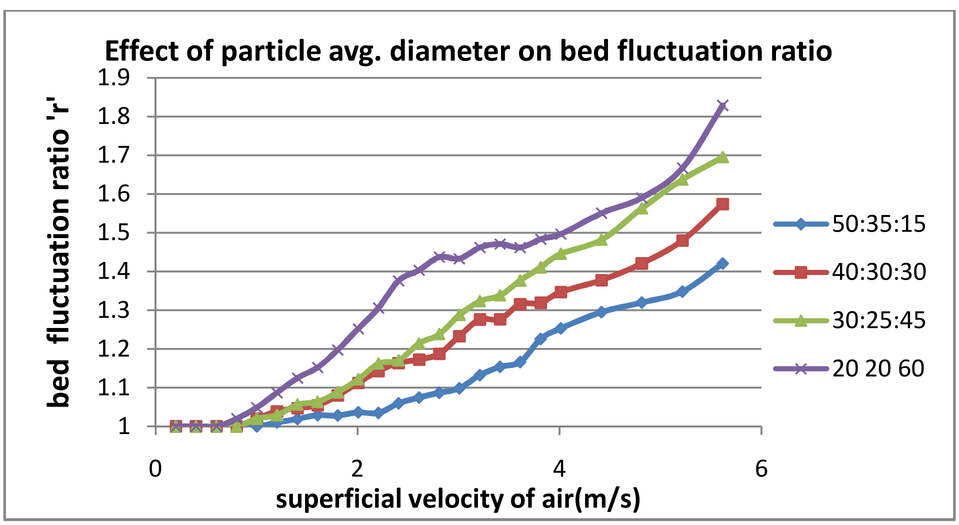

Figure 4.4 Variation of bed fluctuation ratio with gas superficial velocity for cone angle of 7.47° and H_s = 10 cm for different mixtures.

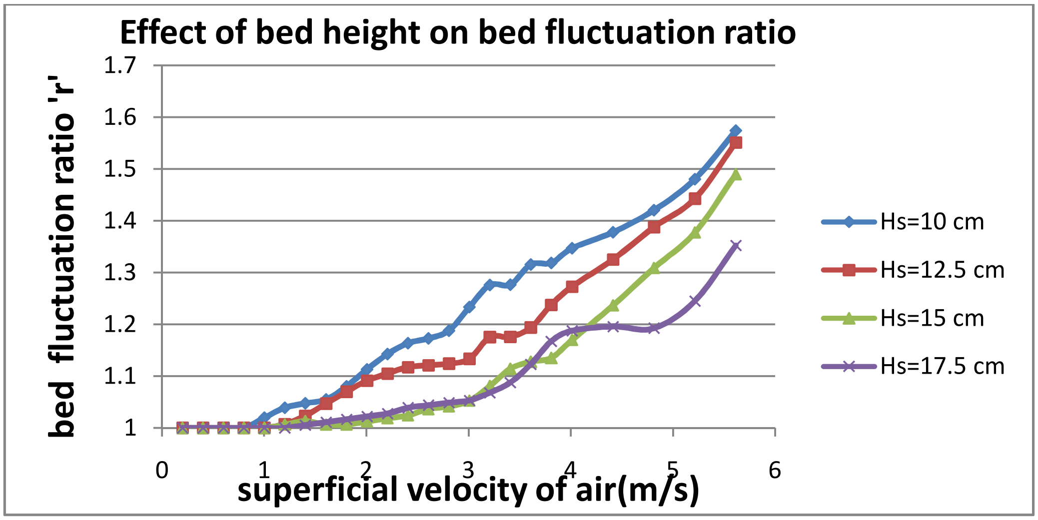

Figure 4.5 Variation of bed fluctuation ratio with gas superficial velocity for cone angle of 7.47° and 40:30:30 mixture for different H_s.

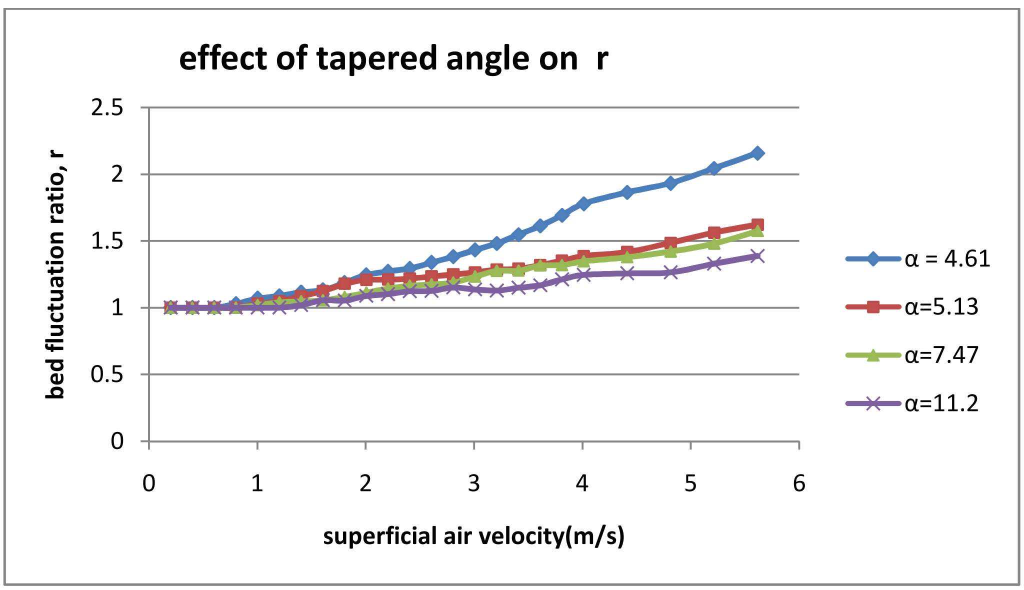

Figure 4.6 Variation of bed fluctuation ratio with gas superficial velocity for different cone angles for 40:30:30 mixture and H_s=10 cm.

(C) **STUDY OF BED EXPANSION RATIO**

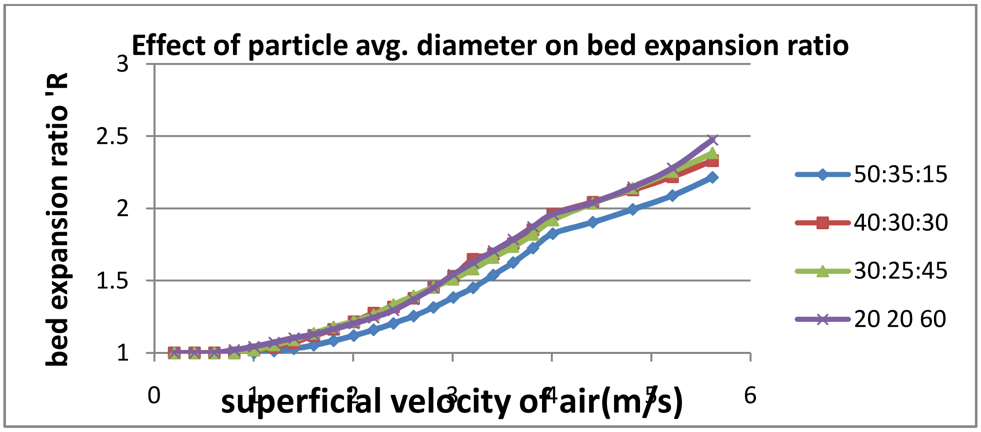

Figure 4.7 Variation of bed expansion ratio with gas superficial velocity for cone angle of 7.47° and H_s = 10 cm for different mixtures.

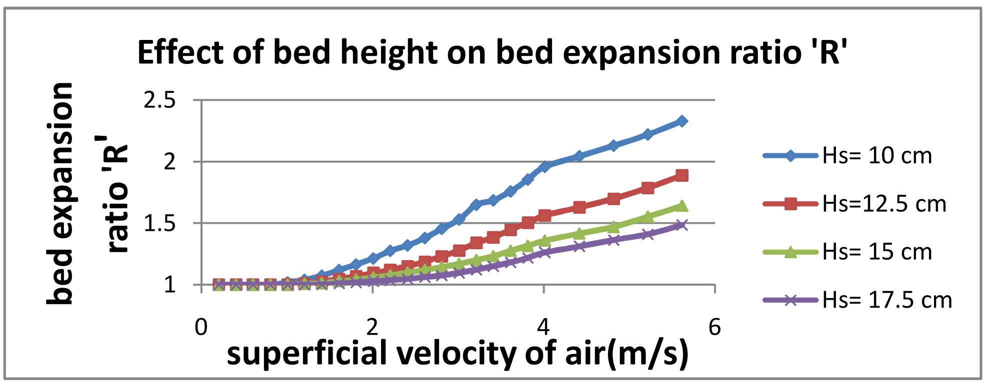

Figure 4.8 Variation of bed expansion ratio with gas superficial velocity for cone angle of 7.47° and 40:30:30 mixture for different H_s.

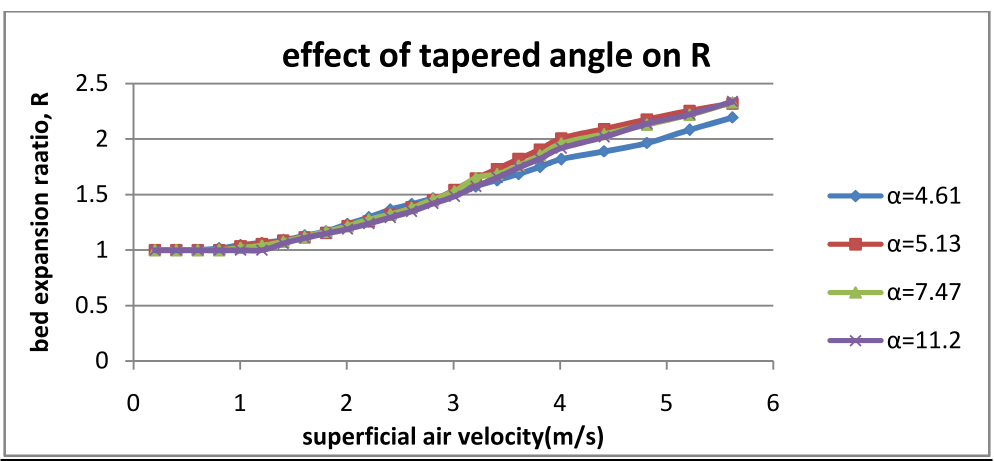

Figure 4.9 Variation of bed expansion ratio with gas superficial velocity for different cone angles for 40:30:30 mixture and H_s=10 cm

DICUSSIONS ON r & R

With increase in particle diameter (i.e. decrease in percentage fines), the bed fluctuation ratio (r) decreases while the bed expansion ratio (R) increases. With increase in bed height both r and R is found to decrease. Also with increase in tapered angle, r decrease and R increases. This is observed for low fine mixture. It may very when the mixture has very high fine content

4.2. DEVELOPMENT OF CORRELATIONS FOR BED FLUCTUATION RATIO BY DIMENSIONAL ANALYSIS

The bed fluctuation and expansion ratio is found to depend on following four dimensionless factors, the exponential power of whose and the constant is obtained from dimensional analysis

$$r \text{ or } R = K\left(\frac{G_f - G_{mf}}{G_{mf}}\right)^a \left(\frac{D_o}{dp_{sm}}\right)^b \left(\frac{h_S}{D_O}\right)^c (\tan\alpha)^d \tag{4.1}$$

TABLE 4.1 Values of the parameters and response for developing correlations

$(G_f-G_{mf})/G_{mf}$	d_o/dp_{sm}	h_s/d_o	tan α	r, exp	R, exp
0.25	35	2.381	0.267	1.019	1.02
1.5	35	2.381	0.267	1.113	1.215
2.75	35	2.381	0.267	1.233	1.65
4	35	2.381	0.267	1.347	1.96
5	35	2.381	0.267	1.42	2.13
3	39.622	2.381	0.267	1.37	1.295
3	37.33	2.381	0.267	1.32	1.58
3	35	2.381	0.267	1.27	1.65
3	32.558	2.381	0.267	1.25	1.825
3	35	2.381	0.267	1.2722	1.65
3	35	2.976	0.267	1.273	1.564
3	35	3.57	0.267	1.209	1.277
3	35	4.167	0.267	1.192	1.365
3	35	2.381	0.1623	1.294	1.365
3	35	2.381	0.181	1.284	1.645
3	35	2.381	0.267	1.27	1.65
3	35	2.381	0.412	1.264	2.14

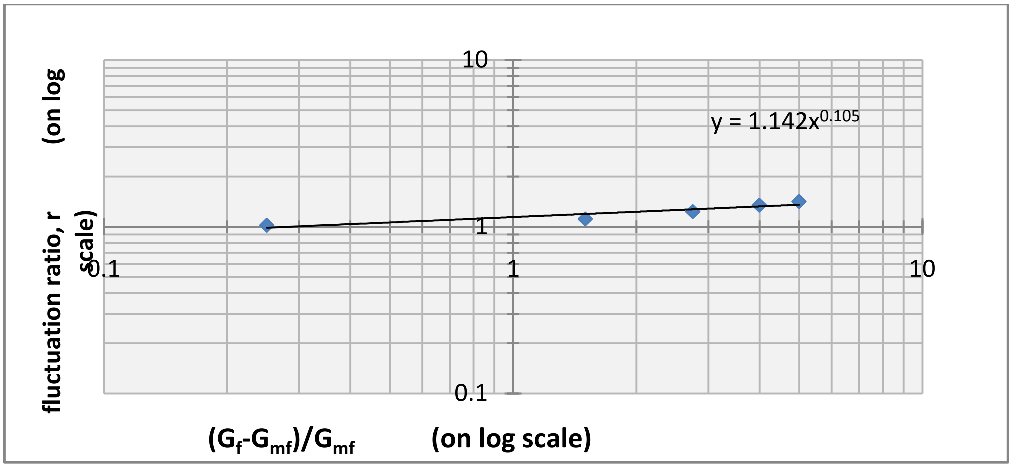

Figure 4.10 Plot of r vs $(G_f-G_{mf})/G_{mf}$ on log-log scale

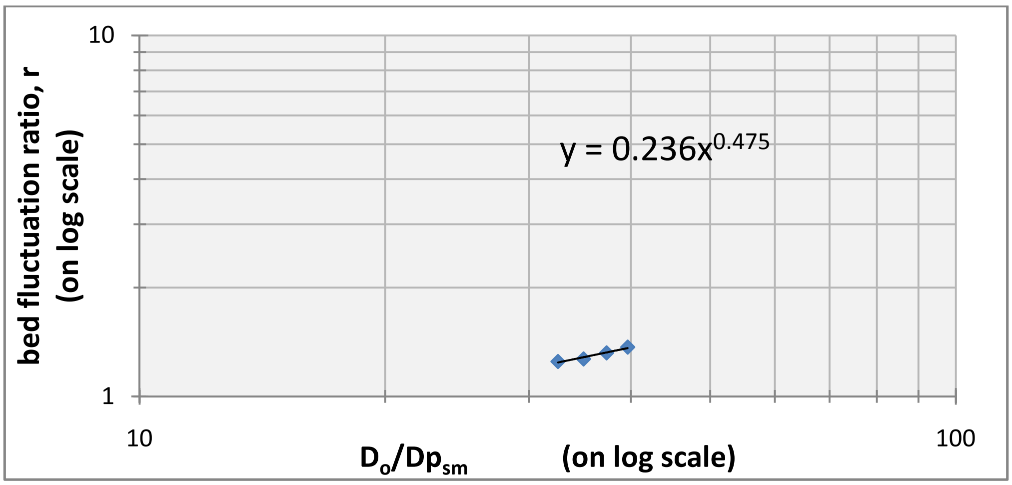

Figure 4.11 Plot of r vs D_o/Dp_{sm} on log-log scale

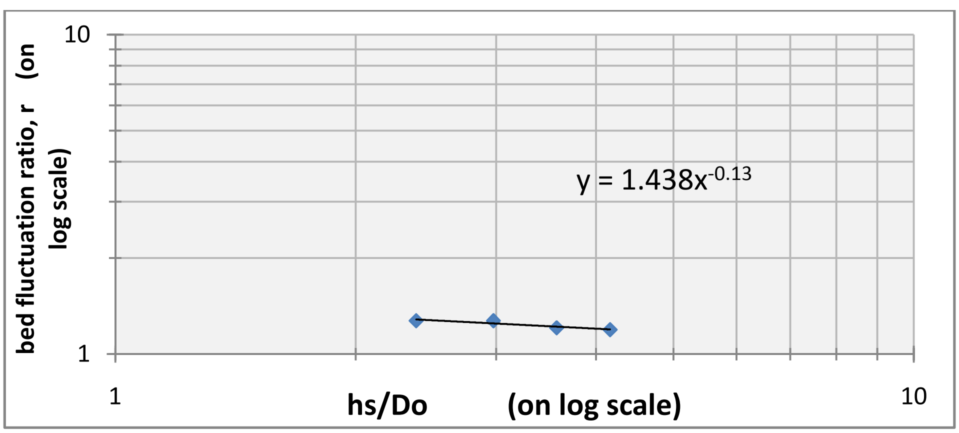

Figure 4.12 Plot of r vs hs/D_o on log-log scale

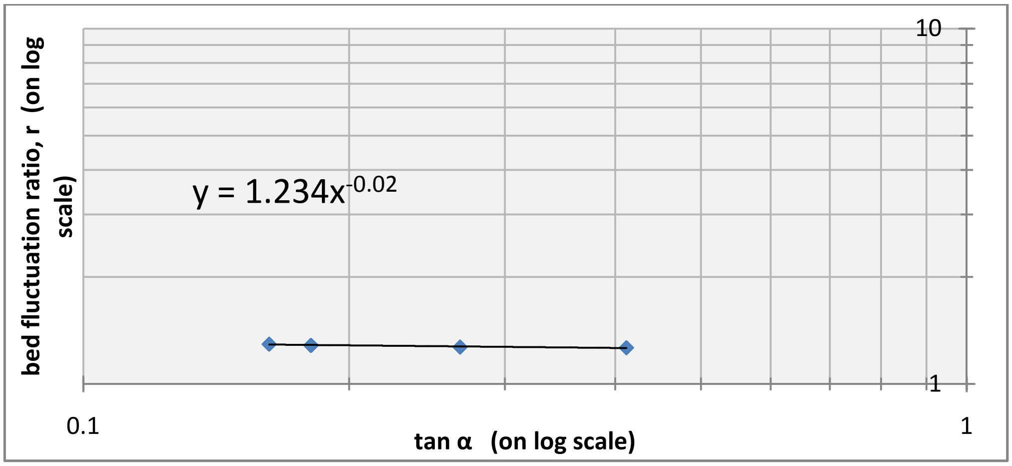

Figure 4.13 Plot of r vs tan α on log-log scale

From earlier graphs, we obtain a=0.105, b=0.475, c= -0.13 and d= -0.02

$$\text{Hence,} product = \left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{0.105} \left(\frac{D_o}{dp_{sm}}\right)^{0.475} \left(\frac{h_S}{D_O}\right)^{-0.13} (\tan\alpha)^{-0.02} \tag{4.2}$$

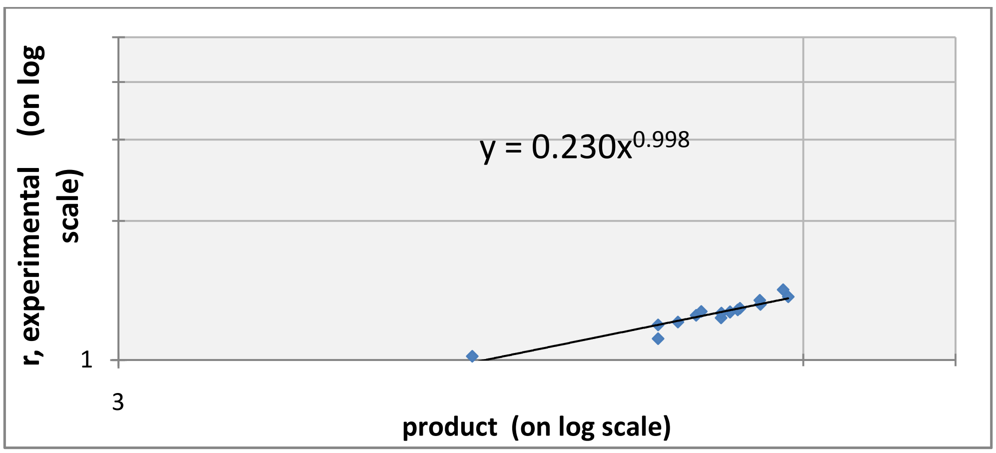

Figure 4.14 Plot of r $_{exp}$ vs product on log-log scale

From above graph, K= 0.230 and n= 0.998

Hence the final correlation is,

$$r = 0.23 \left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{0.1048} \left(\frac{D_o}{dp_{sm}}\right)^{0.478} \left(\frac{h_S}{D_O}\right)^{-0.129} (\tan\alpha)^{-0.0229} \tag{4.3}$$

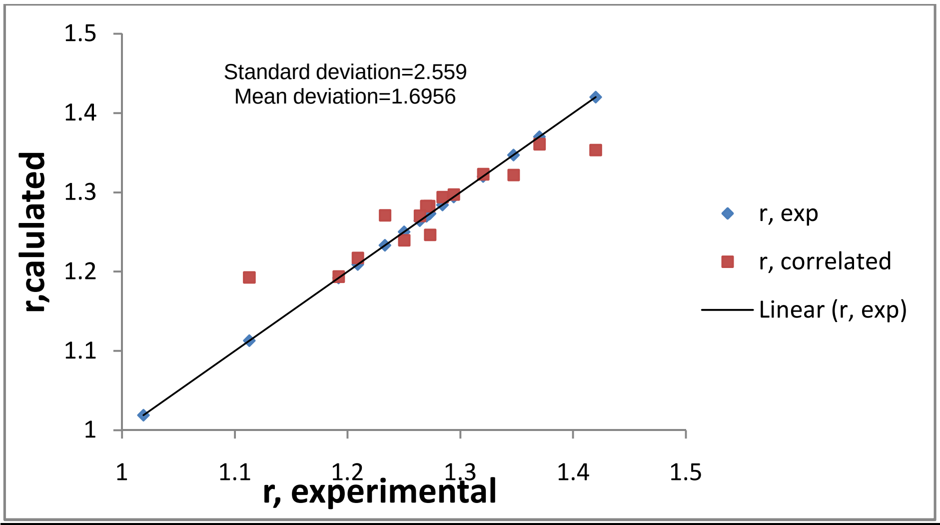

Figure 4.15 Comparison of experimental values of r with correlated value obtained by dimensional analysis

DEVELOPMENT OF CORRELATION FOR BED EXPANSION RATIO

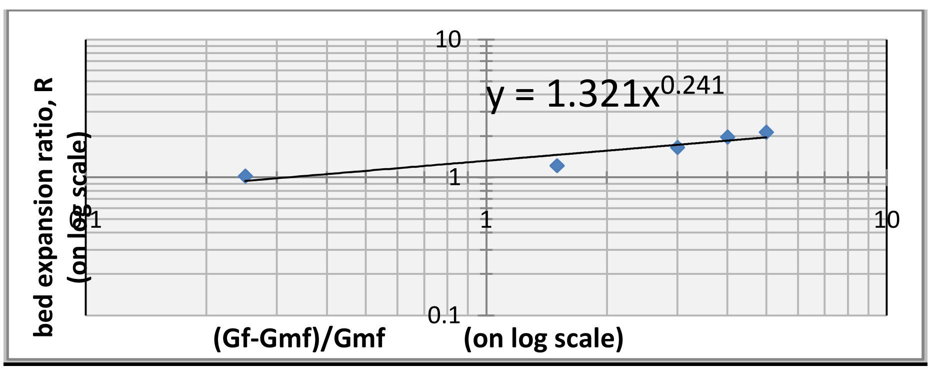

Figure 4.16 Plot of R vs (Gf-Gmf)/Gmf on log-log scale

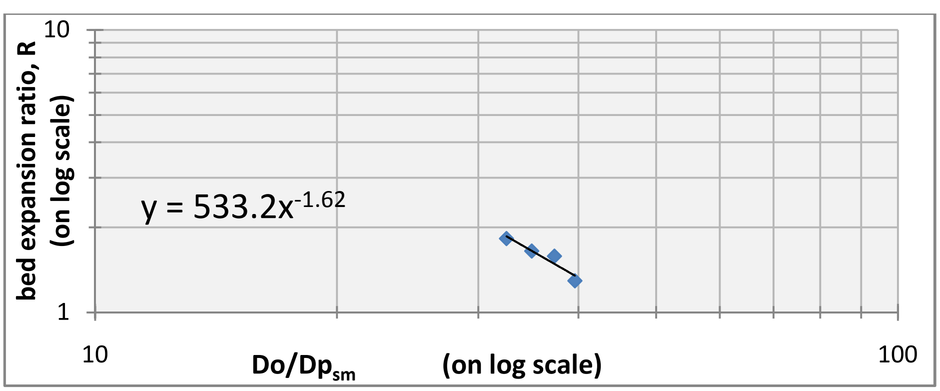

Figure 4.17 Plot of R vs D_o/Dp_{sm} on log-log scale

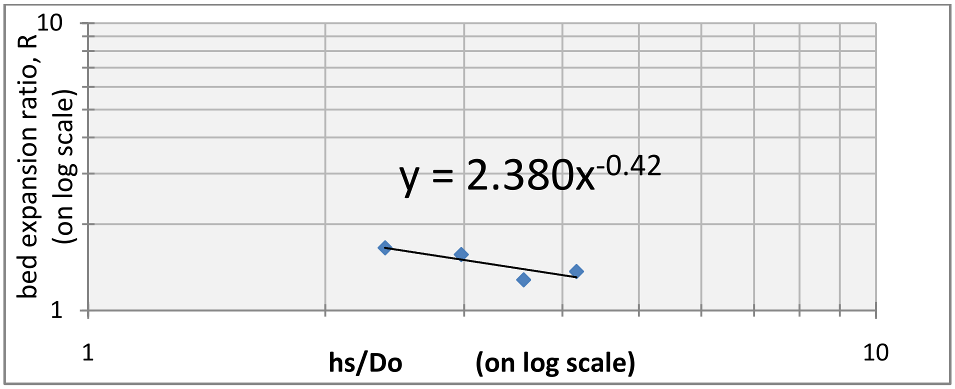

Figure 4.18 Plot of R vs hs/D_o on log-log scale

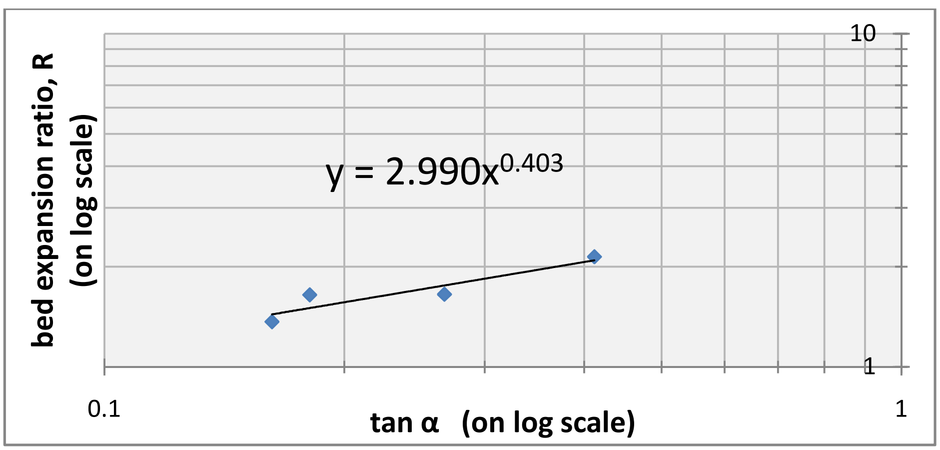

Figure 4.19 Plot of R vs tan α on log-log scale

From above graphs, we obtain a=0.241,b= -1.62,c= -0.42 and d= 0.403

Hence $$product = \left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{0.241} \left(\frac{D_o}{dp_{sm}}\right)^{-1.62} \left(\frac{h_S}{D_O}\right)^{-0.42} (\tan\alpha)^{.403} \tag{4.4}$$

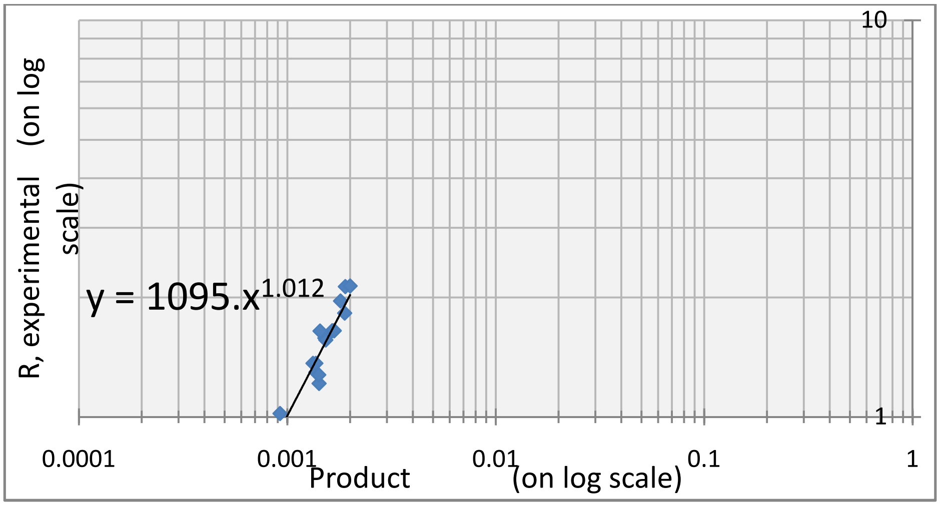

Figure 4.20 Plot of R $_{exp}$ vs product on log-log scale

From above graph,K= 1.095*10^3 and n= 1.012

The final correlation is

$$R = 1.095 * 103 \left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{0.244} \left(\frac{D_o}{dp_{sm}}\right)^{-1.639} \left(\frac{h_S}{D_O}\right)^{-0.425} (\tan\alpha)^{0.408} \tag{4.5}$$

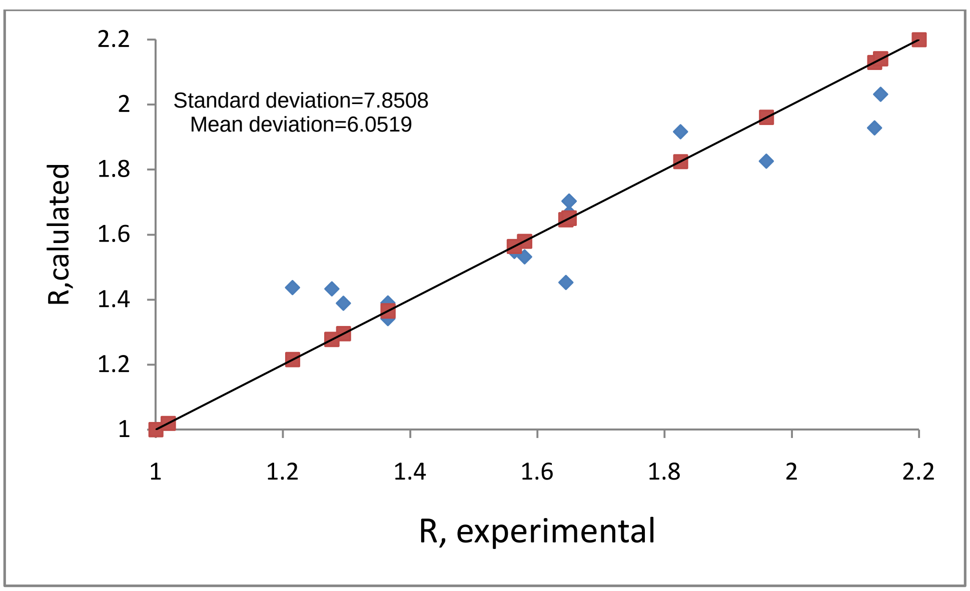

Figure 4.21 Comparison of experimental values of R with correlated value obtained by dimensional analysis

4.3. DEVELOPMENT OF CORRELATIONS FOR BED FLUCTUATION RATIO,BED EXPANSION RATIO and MIN. FLUIDIZATION VELOCITY BY FACTORIAL DESIGN

The method of experimentation is based on statistical design of experiments (Factorial Design Analysis) in order to bring out the interaction effects of variables, which would not be found otherwise by conventional experimentation and to explicitly find out the effect of each of the variables quantitatively on the response. In addition, the number of experiments required is far less compared to the conventional experiments [23]. The scope of the factors consider for factorial experimentation is presented in **table 28**

The variables which affect bed expansion ratios and bed fluctuation ratio in fluidization are static bed height, particle diameter, tapered angle and gas velocity. Thus total numbers of experiments required at two levels for the four variables is sixteen for responses expansion ratio at minimum fluidization velocity.

TABLE 4.2 Scope of the factors for hydrodynamics for Factorial Design Analysis

Variable name	Variable general symbol	Factorial design symbol	Minimum level (-1)	Maximum level (+1)	Magnitude of variables
Excess velocity ratio	(Gf-Gmf)/Gmf	A	0.25	3	0.25 1.5,2.75,3
Inlet diameter to Particle diameter ratio	do/dpsm	B	35	39.62	35,37.33,39.62
Bed height to inlet dia. ratio	Hs/Do	C	2.38	4.167	2.38,2.976,3.57,4.17
Tan(cone angle)	tan α	D	0.267	.412	.267,.412

TABLE 4.3 ANALYSIS OF DATA FOR FACTORIAL DESIGN

Sl. No.	combination	A	B	C	D	r, exp	R, exp
1		0.25	35	2.38	0.267	1.0198	1.02
2	a	3	35	2.38	0.267	1.2759	1.65
3	b	0.25	39.62	2.38	0.267	1.0198	1.02
4	ab	3	39.62	2.38	0.267	1.376	1.295
5	c	0.25	35	4.167	0.267	1.0113	1.0114
6	ac	3	35	4.167	0.267	1.1926	1.3657
7	bc	0.25	39.62	4.167	0.267	1.0057	1.0028
8	abc	3	39.62	4.167	0.267	1.2	1.2571
9	d	0.25	35	2.38	0.412	1.055	1.12
10	ad	3	35	2.38	0.412	1.2645	2.14
11	bd	0.25	39.62	2.38	0.412	1.0098	1.025
12	abd	3	39.62	2.38	0.412	1.271	1.34
13	cd	0.25	35	4.167	0.412	1.00568	1.00857
14	acd	3	35	4.167	0.412	1.183	1.329
15	bcd	0.25	39.62	4.167	0.412	1.00568	1.00857
16	abcd	3	39.62	4.167	0.412	1.0885	1.146

TABLE 4.4 The effects of parameters as per factorial design analysis for r

Sl. No.	treatment combination	r, exp	1	2	3	4	effect= col (4)/8	sum of squares= col(4)^2/16	col(4)/16	% contribution
1	_	1.019	2.2957	4.6915	9.101	17.98	2.2475	20.20502	1.12375	
2	a	1.275	2.395	4.409	8.88	1.71	0.2148	0.184620	0.10742	76.00778
3	b	1.019	2.203	4.600	0.98	-0.03	-0.0039	6.16E-05	-0.002	0.02537
4	ab	1.376	2.205	4.282	0.730	0.07	0.0087	0.000308	0.0043	0.126804
5	c	1.011	2.319	0.612	0.101	-0.66	-0.0829	0.027489	-0.0415	11.31741
6	ac	1.192	2.280	0.375	-0.13	-0.45	-0.0559	0.012504	-0.028	5.148203
7	bc	1.005	2.188	0.470	0.113	-0.15	-0.0188	0.001425	-0.009	0.586694
8	abc	1.2	2.094	0.260	-0.043	-0.23	-0.02915	0.003398	-0.014	1.399313
9	d	1.055	0.256	0.1	-0.282	-0.22	-0.0277	0.002997	-0.013	1.234088
10	ad	1.264	0.356	0.001	-0.38	-0.26	-0.037	0.004131	-0.016	1.700834
11	bd	1.009	0.181	-0.038	-0.237	-0.24	-0.02937	0.003451	-0.014	1.420998
12	abd	1.271	0.194	-0.094	-0.211	-0.16	-0.0194	0.001517	-0.009	0.624587
13	cd	1.005	0.209	0.1	-0.098	-0.1	-0.01242	0.000617	-0.006	0.254232
14	acd	1.183	0.261	0.013	-0.05	0.02	0.00326	4.25E-05	0.0016	0.017528
15	bcd	1.005	0.177	0.051	-0.087	0.04	0.0053	0.000112	0.0026	0.046258
16	abcd	1.088	0.082	-0.094	-0.146	-0.06	-0.0074	0.000219	-0.003	0.090178

The value of the coefficients indicates the magnitude of the effect of the variables and the sign of the coefficient gives the direction of the effect of the variable. That is a positive coefficient indicating an increasing in the value of the responses with increase in the value of the variable and a negative coefficient showing that the response decreases with increase in the value of the variable. The following equation has been obtained,

$$\mathbf{r = 1.12375 + (0.1074*A) - (0.04145*C) - (0.02*A*C)} \qquad (4.6)$$

Similar procedure is followed to find correlations for R and $U_{mf,}$ which are given as,

R = 1.23375+ (0.207*A)-(0.097*B)-(0.0926*C)-(0.073*A*C) (4.7)

U_{mf} =0.8435+ (0.1*C)-(0.0585*D)-(0.1*C*D)-(0.034*B*D) (4.8)

Following comparisons are made between the experimental values and calculated values of U_{mf} and ΔP_{mf} obtained from various correlations already developed for binary regular and single component mixture in tapered fluidized bed.

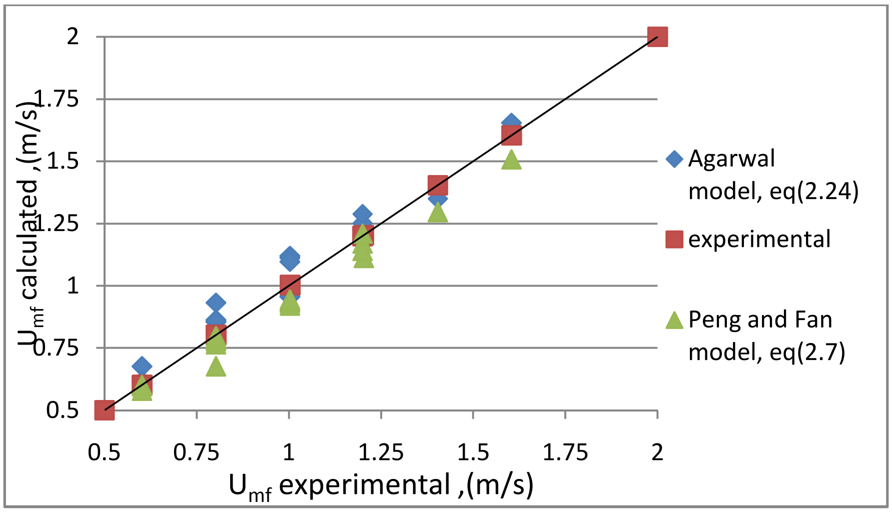

Figure 4.22 Comparison of experimental values with predicted values of U_{mf}

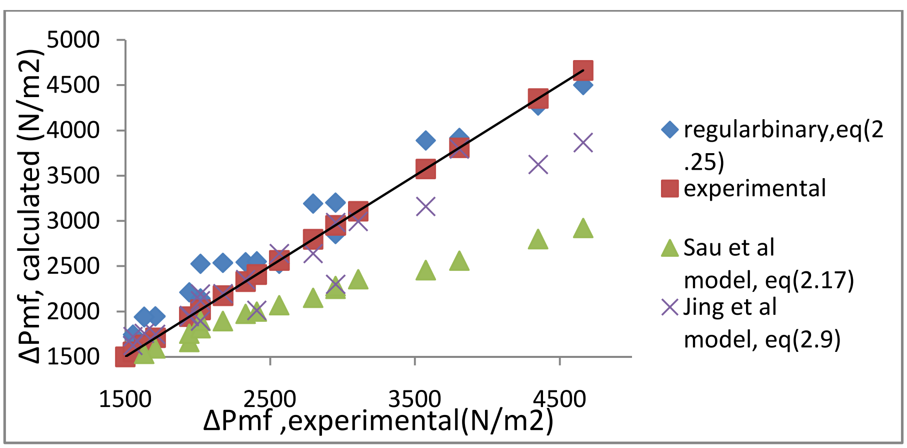

Figure 4.23 Comparison of experimental values with predicted values of ΔPmf

CHAPTER 5

CONCLUSION

CONCLUSION

Based on the above results and discussions, we could conclude that the developed correlations for cylindrical conduits could not be used in case of fluidization in conical conduits. The experimental values of U_{mf} and ΔP_{mf} are found to be generally more than that in binary mixture of irregular particles, as in our case the bed gets fluidized only when the largest particle size of the mixture gets started fluidizing . Moreover it is also seen that bed pressure drop and minimum fluidization velocity is less in case of irregular particles than that in regular particles because of less bed voidage and less channeling and slugging. However there is some deviation seen when compared with the equations with the fluidization of single component mixture. We also found that U_{mf} and ΔP_{mf} is less for tapered beds than that in conventional cylindrical conduits.

Apart from $\Delta P_{mf,}$ the bed fluctuation ratio also decreases by the use of tapered bed, thus preventing the formation and collapse of bubbles by slugging, bubbling that may have resulted in poor gas solid contact and hereby enhancing the quality of fluidization. The bed expansion increase with increase in tapered angle, but then also entrainment of fines is prevented due to gradual change of velocity along the column axis. So due to decrease in pressure drop there is ease of handling, stability of operation, overheat is minimized in case of sensitive products and wide range of superficial gas velocity could be used to fluidize. The above study finds its applicability in optimal designing and operation of fluidized bed reactor that incorporates to determine reactor dimensions, selecting auxiliary equipments and predicting range of major operating variables, particularly in industries like pharmaceuticals, microbiological de-nitrification, bio-mass gasifiers, gasification of coal and catalyst polymerization and gas-solid catalyzed reactions.

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APPENDIX

EXPERIMENTAL OBSERVATIONS

TABLE A.1.: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 10cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	Δ P (in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	3	466.1712	10	10	1	1
2	0.40121005	5.5	854.6472	10	10	1	1
3	0.601815074	8	1243.1232	10	10	1	1
4	0.802420099	11	1709.2944	10	10	1	1
5	1.003025124	10.8	1678.2163	10.3	10.1	1.01980	1.02
6	1.203630149	10.8	1678.2163	10.6	10.2	1.03921	1.04
7	1.404235173	10.7	1662.6772	11	10.5	1.04761	1.075
8	1.604840198	10.8	1678.2163	11.5	10.9	1.05504	1.12
9	1.805445223	10.5	1631.5992	12.1	11.2	1.08035	1.165
10	2.006050248	10.6	1647.1382	12.8	11.5	1.11304	1.215
11	2.206655272	10.6	1647.1382	13.6	11.9	1.14285	1.275
12	2.407260297	10.7	1662.6772	14.2	12.2	1.16393	1.32
13	2.607865322	10.6	1647.1382	14.9	12.7	1.17322	1.38
14	2.808470347	10.7	1662.6772	15.8	13.3	1.18797	1.455
15	3.009075371	10.7	1662.6772	16.9	13.7	1.23357	1.53
16	3.209680396	10.8	1678.2163	18.5	14.5	1.27586	1.65
17	3.410285421	10.9	1693.7553	18.9	14.8	1.27702	1.685
18	3.610890446	10.9	1693.7553	20	15.2	1.31578	1.76
19	3.81149547	11	1709.2944	21.1	16	1.31875	1.855
20	4.012100495	10.9	1693.7553	22.5	16.7	1.34730	1.96
22	4.413310545	11	1709.2944	23.7	17.2	1.37790	2.045
24	4.814520594	11	1709.2944	25	17.6	1.42045	2.13
26	5.215730644	11	1709.2944	26.5	17.9	1.48044	2.22
28	5.616940693	11	1709.2944	28.5	18.1	1.57458	2.33

TABLE A.2: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 12.5cm and α= 7.47°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP(cm)**	**Δ P (in N/m^2)**	**h1(max) cm**	**h2(min) , cm**	**r**	**R**
1	0.200605025	4	621.5616	12.5	12.5	1	1
2	0.40121005	6.5	1010.0376	12.5	12.5	1	1
3	0.601815074	9	1398.5136	12.5	12.5	1	1
4	0.802420099	11	1709.2944	12.5	12.5	1	1
5	1.003025124	14	2175.4656	12.5	12.5	1	1
6	1.203630149	13.9	2159.9265	12.7	12.6	1.0079	1.012
7	1.404235173	13.8	2144.3875	13	12.7	1.0236	1.028
8	1.604840198	13.8	2144.3875	13.4	12.8	1.0468	1.048
9	1.805445223	13.7	2128.8484	13.8	12.9	1.0697	1.068
10	2.006050248	13.6	2113.3094	14.3	13.1	1.0916	1.096
11	2.206655272	13.6	2113.3094	14.7	13.3	1.1052	1.12
12	2.407260297	13.6	2113.3094	15.2	13.6	1.1176	1.152
13	2.607865322	13.7	2128.8484	15.7	14	1.1214	1.188
14	2.808470347	13.6	2113.3094	16.3	14.5	1.1241	1.232
15	3.009075371	13.7	2128.8484	17	15	1.1333	1.28
16	3.209680396	13.7	2128.8484	18.1	15.4	1.1753	1.34
17	3.410285421	13.8	2144.3875	18.7	15.9	1.1761	1.384
18	3.610890446	13.8	2144.3875	19.7	16.5	1.1939	1.448
19	3.81149547	13.9	2159.9265	20.8	16.8	1.2380	1.504
20	4.012100495	13.8	2144.3875	21.9	17.2	1.2732	1.564
22	4.413310545	13.9	2159.9265	23.2	17.5	1.3257	1.628
24	4.814520594	13.9	2159.9265	24.7	17.8	1.387	1.7
26	5.215730644	14	2175.4656	26.4	18.3	1.4426	1.788
28	5.616940693	14	2175.4656	28.7	18.5	1.5513	1.888

TABLE A.3: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 15cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	Δ P (in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	5	776.952	15	15	1	1
2	0.40121005	7	1087.7328	15	15	1	1
3	0.601815074	9.5	1476.2088	15	15	1	1
4	0.802420099	15	2330.856	15	15	1	1
5	1.003025124	18	2797.0272	15	15	1	1
6	1.203630149	17.3	2688.25392	15.2	15.1	1.00662	1.01
7	1.404235173	17.3	2688.25392	15.3	15.1	1.01324	1.01
8	1.604840198	17.2	2672.71488	15.5	15.4	1.00649	1.03
9	1.805445223	17.4	2703.79296	15.7	15.6	1.00641	1.04
10	2.006050248	17.3	2688.25392	16	15.8	1.01265	1.06
11	2.206655272	17.4	2703.79296	16.3	16	1.01875	1.07
12	2.407260297	17.3	2688.25392	16.7	16.3	1.02454	1.1
13	2.607865322	17.2	2672.71488	17.1	16.5	1.03636	1.12
14	2.808470347	17.2	2672.71488	17.5	16.8	1.04166	1.14
15	3.009075371	17.1	2657.17584	18	17.1	1.05263	1.17
16	3.209680396	17.3	2688.25392	18.7	17.3	1.08092	1.2
17	3.410285421	17.3	2688.25392	19.5	17.5	1.11428	1.23
18	3.610890446	17.5	2719.332	20.3	18	1.12777	1.27
19	3.81149547	17.6	2734.87104	21	18.5	1.13513	1.31
20	4.012100495	17.6	2734.87104	22	18.8	1.17021	1.36
22	4.413310545	17.7	2750.41008	23.5	19	1.23684	1.41
24	4.814520594	17.6	2734.87104	25	19.1	1.30890	1.47
26	5.215730644	17.6	2734.87104	27	19.6	1.37755	1.55
28	5.616940693	17.6	2734.87104	29.5	19.8	1.48989	1.64

TABLE A.4: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 17.5cm and α= 7.47°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP,(in cm)**	**Δ P (in N/m^2)**	**h1(max), cm**	**h2(min), cm**	**r**	**R**
1	0.200605025	6	932.3424	17.5	17.5	1	1
2	0.40121005	9.5	1476.2088	17.5	17.5	1	1
3	0.601815074	12.4	1926.84096	17.5	17.5	1	1
4	0.802420099	15.3	2377.47312	17.5	17.5	1	1
5	1.003025124	18.5	2874.7224	17.5	17.5	1	1
6	1.203630149	23	3573.9792	17.5	17.5	1	1
7	1.404235173	21.9	3403.04976	17.6	17.5	1.0057	1.0029
8	1.604840198	21.7	3371.97168	17.8	17.6	1.0114	1.0114
9	1.805445223	21.6	3356.43264	18	17.7	1.0169	1.02
10	2.006050248	21.6	3356.43264	18.2	17.8	1.0225	1.0286
11	2.206655272	21.5	3340.8936	18.4	17.9	1.0279	1.0371
12	2.407260297	21.6	3356.43264	18.7	18	1.0389	1.0486
13	2.607865322	21.5	3340.8936	19	18.2	1.044	1.0629
14	2.808470347	21.6	3356.43264	19.3	18.4	1.0489	1.0771
15	3.009075371	21.8	3387.51072	19.7	18.7	1.0535	1.0971
16	3.209680396	21.8	3387.51072	20.3	19	1.0684	1.1229
17	3.410285421	22	3418.5888	21	19.3	1.0881	1.1514
18	3.610890446	22.3	3465.20592	21.9	19.5	1.1231	1.1829
19	3.81149547	22.3	3465.20592	23	19.7	1.1675	1.22
20	4.012100495	22.5	3496.284	24	20.2	1.1881	1.2629
22	4.413310545	22.6	3511.82304	25	20.9	1.1962	1.3114
24	4.814520594	22.6	3511.82304	26	21.8	1.1927	1.3657
26	5.215730644	22.6	3511.82304	27.4	22	1.2455	1.4114
28	5.616940693	22.6	3511.82304	29.9	22.1	1.3529	1.4857

TABLE A.5: Variation of pressure drop with superficial gas velocity for 30:25:45 mixture with h_s= 10 cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	Δ P (in N/m^2)	h1(max)	h2(min)	r	R
1	0.200605025	2	310.7808	10	10	1	1
2	0.40121005	4.5	699.2568	10	10	1	1
3	0.601815074	7.1	1103.27184	10	10	1	1
4	0.802420099	10.5	1631.5992	10	10	1	1
5	1.003025124	9.9	1538.36496	10.3	10.1	1.0198	1.02
6	1.203630149	10	1553.904	10.7	10.4	1.0288	1.055
7	1.404235173	9.8	1522.82592	11.2	10.6	1.0566	1.09
8	1.604840198	9.7	1507.28688	11.7	11	1.0636	1.135
9	1.805445223	9.8	1522.82592	12.3	11.3	1.0885	1.18
10	2.006050248	9.8	1522.82592	12.9	11.5	1.1217	1.22
11	2.206655272	9.9	1538.36496	13.6	11.7	1.1624	1.265
12	2.407260297	9.7	1507.28688	14.4	12.3	1.1707	1.335
13	2.607865322	9.6	1491.74784	15.3	12.6	1.2143	1.395
14	2.808470347	9.8	1522.82592	16.1	13	1.2385	1.455
15	3.009075371	9.8	1522.82592	17	13.2	1.2879	1.51
16	3.209680396	9.7	1507.28688	18	13.6	1.3235	1.58
17	3.410285421	9.7	1507.28688	19	14.2	1.338	1.66
18	3.610890446	9.7	1507.28688	20.1	14.6	1.3767	1.735
19	3.81149547	9.8	1522.82592	21.3	15.1	1.4106	1.82
20	4.012100495	9.8	1522.82592	22.7	15.7	1.4459	1.92
22	4.413310545	9.9	1538.36496	24.3	16.4	1.4817	2.035
24	4.814520594	9.9	1538.36496	26.1	16.7	1.5629	2.14
26	5.215730644	9.9	1538.36496	28	17.1	1.6374	2.255
28	5.616940693	9.9	1538.36496	30	17.7	1.6949	2.385

TABLE A.6: Variation of pressure drop with superficial gas velocity for 30:25:45 mixture with h_s= 12.5 cm and α= 7.47°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP(cm)**	**Δ P (in N/m^2)**	**h1(max), cm**	**h2(min), cm**	**r**	**R**
1	0.20060502	3	466.1712	12.5	12.5	1	1
2	0.40121005	5.6	870.18624	12.5	12.5	1	1
3	0.60181507	8.7	1351.89648	12.5	12.5	1	1
4	0.8024201	13	2020.0752	12.5	12.5	1	1
5	1.00302512	12.2	1895.76288	12.7	12.6	1.007937	1.012
6	1.20363015	12	1864.6848	12.9	12.6	1.02381	1.02
7	1.40423517	12	1864.6848	13.3	12.9	1.031008	1.048
8	1.6048402	11.9	1849.14576	13.7	13.5	1.014815	1.088
9	1.80544522	11.7	1818.06768	14.2	13.5	1.051852	1.108
10	2.00605025	11.8	1833.60672	14.7	13.8	1.065217	1.14
11	2.20665527	11.8	1833.60672	15.2	14.2	1.070423	1.176
12	2.4072603	11.7	1818.06768	15.8	14.7	1.07483	1.22
13	2.60786532	11.8	1833.60672	16.4	15	1.093333	1.256
14	2.80847035	11.9	1849.14576	17.1	15.6	1.096154	1.308
15	3.00907537	11.9	1849.14576	17.8	16.3	1.092025	1.364
16	3.2096804	11.8	1833.60672	18.6	16.6	1.120482	1.408
17	3.41028542	11.9	1849.14576	19.7	17	1.158824	1.468
18	3.61089045	12	1864.6848	21.2	17	1.247059	1.528
19	3.81149547	12	1864.6848	22.5	17.2	1.30814	1.588
20	4.0121005	11.9	1849.14576	23.6	17.4	1.356322	1.64
22	4.41331054	12	1864.6848	24.8	17.5	1.417143	1.692
24	4.81452059	11.9	1849.14576	26.2	18	1.455556	1.768
26	5.21573064	12	1864.6848	28	19.1	1.465969	1.884
28	5.61694069	12	1864.6848	30.9	19.8	1.560606	2.028

TABLE A.7: Variation of pressure drop with superficial gas velocity for 30:25:45 mixture with h_s= 15 cm and α= 7.47°

flow rate(m^3/hr)	gasvelocity(in m/s)	ΔP(cm)	Δ P (in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	4	621.5616	15	15	1	1
2	0.40121005	6.7	1041.11568	15	15	1	1
3	0.601815074	9.2	1429.59168	15	15	1	1
4	0.802420099	13.2	2051.15328	15	15	1	1
5	1.003025124	16.5	2563.9416	15	15	1	1
6	1.203630149	15	2330.856	15.2	15.1	1.0066	1.01
7	1.404235173	14.5	2253.1608	15.4	15.2	1.0132	1.02
8	1.604840198	14.7	2284.23888	15.6	15.4	1.013	1.033333
9	1.805445223	14.4	2237.62176	15.8	15.6	1.0128	1.046667
10	2.006050248	14.5	2253.1608	16.1	15.8	1.019	1.063333
11	2.206655272	14.7	2284.23888	16.5	16.1	1.0248	1.086667
12	2.407260297	14.7	2284.23888	17	16.4	1.0366	1.113333
13	2.607865322	14.6	2268.69984	17.7	16.7	1.0599	1.146667
14	2.808470347	14.5	2253.1608	18.5	17.1	1.0819	1.186667
15	3.009075371	14.7	2284.23888	19.5	17.5	1.1143	1.233333
16	3.209680396	14.6	2268.69984	20.6	17.8	1.1573	1.28
17	3.410285421	14.9	2315.31696	21.6	18.4	1.1739	1.333333
18	3.610890446	14.9	2315.31696	22.7	18.9	1.2011	1.386667
19	3.81149547	14.8	2299.77792	23.9	19.5	1.2256	1.446667
20	4.012100495	15	2330.856	25.3	20.2	1.2525	1.516667
22	4.413310545	15.2	2361.93408	26.5	20.7	1.2802	1.573333
24	4.814520594	15.2	2361.93408	27.9	21.3	1.3099	1.64
26	5.215730644	15.2	2361.93408	28.8	21.8	1.3211	1.686667
28	5.616940693	15.2	2361.93408	31.9	22	1.45	1.796667

TABLE A.8: Variation of pressure drop with superficial gas velocity for 30:25:45 mixture with h_s= 17.5 cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	ΔP(in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605	5	776.952	17.5	17.5	1	1
2	0.40121	8.1	1258.66224	17.5	17.5	1	1
3	0.6018151	10.9	1693.75536	17.5	17.5	1	1
4	0.8024201	13.8	2144.38752	17.5	17.5	1	1
5	1.0030251	16.9	2626.09776	17.5	17.5	1	1
6	1.2036301	20	3107.808	17.5	17.5	1	1
7	1.4042352	18.2	2828.10528	17.7	17.6	1.005682	1.008571
8	1.6048402	17.9	2781.48816	17.9	17.7	1.011299	1.017143
9	1.8054452	17.9	2781.48816	18.1	17.9	1.011173	1.028571
10	2.0060502	18.2	2828.10528	18.4	18	1.022222	1.04
11	2.2066553	17.7	2750.41008	18.7	18.2	1.027473	1.054286
12	2.4072603	17.8	2765.94912	19	18.5	1.027027	1.071429
13	2.6078653	17.7	2750.41008	19.3	18.6	1.037634	1.082857
14	2.8084703	17.5	2719.332	19.7	18.8	1.047872	1.1
15	3.0090754	17.4	2703.79296	20.4	19	1.073684	1.125714
16	3.2096804	17.7	2750.41008	21	19.2	1.09375	1.148571
17	3.4102854	17.8	2765.94912	21.8	19.7	1.106599	1.185714
18	3.6108904	17.7	2750.41008	22.8	20	1.14	1.222857
19	3.8114955	17.8	2765.94912	24	20.4	1.176471	1.268571
20	4.0121005	17.9	2781.48816	25.3	20.5	1.234146	1.308571
22	4.4133105	17.8	2765.94912	26.6	20.8	1.278846	1.354286
24	4.8145206	17.9	2781.48816	28	21	1.333333	1.4
26	5.2157306	17.9	2781.48816	29.4	21.3	1.380282	1.448571
28	5.6169407	17.9	2781.48816	30.9	21.5	1.437209	1.497143

TABLE A.9: Variation of pressure drop with superficial gas velocity for 50:35:15 mixture with h_s= 10 cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	ΔP(in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	4	621.5616	10	10	1	1
2	0.40121005	6.5	1010.0376	10	10	1	1
3	0.601815074	8	1243.1232	10	10	1	1
4	0.802420099	11	1709.2944	10	10	1	1
5	1.003025124	12.5	1942.38	10	10	1	1
6	1.203630149	12	1864.6848	10.2	10.1	1.00990	1.015
7	1.404235173	12	1864.6848	10.4	10.2	1.01960	1.03
8	1.604840198	11.8	1833.60672	10.7	10.4	1.02884	1.055
9	1.805445223	11.7	1818.06768	11	10.7	1.02803	1.085
10	2.006050248	11.8	1833.60672	11.4	11	1.03636	1.12
11	2.206655272	11.8	1833.60672	11.8	11.4	1.03508	1.16
12	2.407260297	11.7	1818.06768	12.4	11.7	1.059829	1.205
13	2.607865322	11.6	1802.52864	13	12.1	1.07438	1.255
14	2.808470347	11.6	1802.52864	13.7	12.6	1.08730	1.315
15	3.009075371	11.8	1833.60672	14.5	13.2	1.09848	1.385
16	3.209680396	11.8	1833.60672	15.4	13.6	1.13235	1.45
17	3.410285421	12	1864.6848	16.5	14.3	1.15384	1.54
18	3.610890446	11.9	1849.14576	17.5	15	1.16666	1.625
19	3.81149547	11.9	1849.14576	19	15.5	1.22580	1.725
20	4.012100495	12	1864.6848	20.3	16.2	1.25308	1.825
22	4.413310545	12	1864.6848	21.5	16.6	1.29518	1.905
24	4.814520594	12	1864.6848	22.7	17.2	1.31976	1.995
26	5.215730644	12	1864.6848	24	17.8	1.34831	2.09
28	5.616940693	12.1	1880.22384	26	18.3	1.42076	2.215

TABLE A.10: Variation of pressure drop with superficial gas velocity for 50:35:15 mixture with h_s= 12.5 cm and α= 7.47°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP(cm)**	**Δ P (in N/m^2)**	**h1(max), cm**	**h2(min), cm**	**r**	**R**
1	0.200605025	5	776.952	12.5	12.5	1	1
2	0.40121005	7	1087.7328	12.5	12.5	1	1
3	0.601815074	8.7	1351.89648	12.5	12.5	1	1
4	0.802420099	12	1864.6848	12.5	12.5	1	1
5	1.003025124	12.8	1988.99712	12.5	12.5	1	1
6	1.203630149	15	2330.856	12.5	12.5	1	1
7	1.404235173	14.6	2268.69984	12.7	12.6	1.007937	1.012
8	1.604840198	14.4	2237.62176	12.9	12.8	1.007813	1.028
9	1.805445223	14.1	2191.00464	13.1	13	1.007692	1.044
10	2.006050248	14.3	2222.08272	13.4	13.2	1.015152	1.064
11	2.206655272	14.2	2206.54368	13.8	13.5	1.022222	1.092
12	2.407260297	14.3	2222.08272	14.2	13.8	1.028986	1.12
13	2.607865322	14.3	2222.08272	14.7	14.1	1.042553	1.152
14	2.808470347	14.5	2253.1608	15.3	14.3	1.06993	1.184
15	3.009075371	14.5	2253.1608	15.9	14.6	1.089041	1.22
16	3.209680396	14.6	2268.69984	16.7	15	1.113333	1.268
17	3.410285421	14.7	2284.23888	17.6	15.5	1.135484	1.324
18	3.610890446	14.8	2299.77792	18.7	16	1.16875	1.388
19	3.81149547	14.7	2284.23888	19.9	16.7	1.191617	1.464
20	4.012100495	14.7	2284.23888	21	17.5	1.2	1.54
22	4.413310545	14.8	2299.77792	22.2	18	1.233333	1.608
24	4.814520594	14.7	2284.23888	23.5	18.8	1.25	1.692
26	5.215730644	14.7	2284.23888	25	19.6	1.27551	1.784
28	5.616940693	14.7	2284.23888	26.9	20.2	1.331683	1.884

TABLE A.11: Variation of pressure drop with superficial gas velocity for 50:35:15 mixture with h_s= 15 cm and α= 7.47°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP(cm)**	**Δ P (in N/m^2)**	**h1(max) , cm**	**h2(hmin) , cm**	**r**	**R**
1	0.200605025	6	932.3424	15	15	1	1
2	0.40121005	8	1243.1232	15	15	1	1
3	0.601815074	9.8	1522.8259	15	15	1	1
4	0.802420099	12.6	1957.9190	15	15	1	1
5	1.003025124	15.3	2377.4731	15	15	1	1
6	1.203630149	17.6	2734.8710	15	15	1	1
7	1.404235173	19	2952.4176	15	15	1	1
8	1.604840198	18.8	2921.3395	15.2	15.1	1.00662	1.01
9	1.805445223	18.6	2890.2614	15.4	15.2	1.01315	1.02
10	2.006050248	18.4	2859.1833	15.6	15.4	1.01298	1.03333
11	2.206655272	18.5	2874.7224	16	15.7	1.01910	1.05666
12	2.407260297	18.5	2874.7224	16.3	15.9	1.02515	1.07333
13	2.607865322	18.6	2890.2614	16.6	16	1.0375	1.08666
14	2.808470347	18.6	2890.2614	17	16.2	1.04938	1.10666
15	3.009075371	18.6	2890.2614	17.3	16.4	1.05487	1.12333
16	3.209680396	18.7	2905.8004	17.7	16.7	1.05988	1.14666
17	3.410285421	18.7	2905.8004	18.2	17	1.07058	1.17333
18	3.610890446	18.6	2890.2614	18.8	17.2	1.09302	1.2
19	3.81149547	18.6	2890.2614	19.5	17.4	1.12069	1.23
20	4.012100495	18.8	2921.3395	20.4	17.7	1.15254	1.27
22	4.413310545	18.8	2921.3395	21.4	18	1.18888	1.31333
24	4.814520594	18.8	2921.3395	22.5	18.2	1.23626	1.35666
26	5.215730644	18.7	2905.8004	23.6	18.4	1.28260	1.4
28	5.616940693	18.7	2905.8004	25	18.7	1.33689	1.45666

TABLE A.12: Variation of pressure drop with superficial gas velocity for 50:35:15 mixture with h_s= 17.5 cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	Δ P (in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	7	1087.7328	17.5	17.5	1	1
2	0.40121005	8.6	1336.3574	17.5	17.5	1	1
3	0.601815074	10.4	1616.0601	17.5	17.5	1	1
4	0.802420099	13	2020.0752	17.5	17.5	1	1
5	1.003025124	15.8	2455.1683	17.5	17.5	1	1
6	1.203630149	18.2	2828.1052	17.5	17.5	1	1
7	1.404235173	21.3	3309.8155	17.5	17.5	1	1
8	1.604840198	24.5	3807.0648	17.5	17.5	1	1
9	1.805445223	24	3729.3696	17.7	17.6	1.0056	1.0085
10	2.006050248	23.8	3698.2915	17.9	17.7	1.0112	1.0171
11	2.206655272	23.7	3682.7524	18.1	17.8	1.0168	1.0257
12	2.407260297	23.6	3667.2134	18.3	17.8	1.0280	1.0314
13	2.607865322	23.7	3682.752	18.6	18	1.0333	1.0457
14	2.808470347	23.7	3682.7524	18.9	18.1	1.0441	1.0571
15	3.009075371	23.6	3667.2134	19.2	18.3	1.0491	1.0714
16	3.209680396	23.8	3698.2915	19.7	18.4	1.0706	1.0885
17	3.410285421	23.8	3698.2915	20	18.7	1.0695	1.1057
18	3.610890446	24	3729.3696	20.8	19	1.0947	1.1371
19	3.81149547	23.8	3698.2915	21.4	19.5	1.0974	1.1685
20	4.012100495	24.1	3744.9086	22.1	19.9	1.1105	1.2
22	4.413310545	24.1	3744.9086	22.9	20.5	1.1170	1.24
24	4.814520594	24.2	3760.4476	23.8	21.1	1.1279	1.2828
26	5.215730644	24.1	3744.9086	24.8	21.8	1.1376	1.3314
28	5.616940693	24.1	3744.9086	26	22.5	1.1555	1.3857
30	6.018150743	24	3729.3696	28	23.7	1.1814	1.4771

TABLE A.13: Variation of pressure drop with superficial gas velocity for 20:20:60 mixture with h_s= 10cm and α= 7.47°

flow rate(m^3/hr)	gasvelocity(in m/s)	ΔP(cm)	ΔP(in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	3	466.1712	10	10	1	1
2	0.40121005	6.5	1010.0376	10	10	1	1
3	0.601815074	10	1553.904	10	10	1	1
4	0.802420099	9.9	1538.36496	10.3	10.1	1.019802	1.02
5	1.003025124	9.8	1522.82592	10.7	10.2	1.04902	1.045
6	1.203630149	9.7	1507.28688	11.2	10.3	1.087379	1.075
7	1.404235173	9.6	1491.74784	11.7	10.4	1.125	1.105
8	1.604840198	9.5	1476.2088	12.1	10.5	1.152381	1.13
9	1.805445223	9.4	1460.66976	12.7	10.6	1.198113	1.165
10	2.006050248	9.4	1460.66976	13.4	10.7	1.252336	1.205
11	2.206655272	9.5	1476.2088	14.1	10.8	1.305556	1.245
12	2.407260297	9.7	1507.28688	15	10.9	1.376147	1.295
13	2.607865322	9.9	1538.36496	16	11.4	1.403509	1.37
14	2.808470347	9.9	1538.36496	17.1	11.9	1.436975	1.45
15	3.009075371	10	1553.904	18.2	12.7	1.433071	1.545
16	3.209680396	9.9	1538.36496	19.3	13.2	1.462121	1.625
17	3.410285421	9.8	1522.82592	20.3	13.8	1.471014	1.705
18	3.610890446	10	1553.904	21.2	14.5	1.46206	1.785
19	3.81149547	9.9	1538.36496	22.4	15.1	1.48344	1.875
20	4.012100495	10	1553.904	23.5	15.7	1.49681	1.96
22	4.413310545	10	1553.904	24.8	16	1.55	2.04
24	4.814520594	10	1553.904	26.4	16.6	1.59036	2.15
26	5.215730644	10	1553.904	28.5	17.1	1.666667	2.28
28	5.616940693	10	1553.904	32	17.5	1.82857	2.475

TABLE A.14: Variation of pressure drop with superficial gas velocity for 20:20:60 mixture with h_s= 12.5cm and α= 7.47°

flow rate(m^3/hr)	**gas vel.(in m/s)**	**ΔP(cm)**	**Δ P (in N/m^2)**	**h1(max), cm**	**h2(min), cm**	**r**	**R**
1	0.200605025	4.5	699.2568	12.5	12.5	1	1
2	0.40121005	7.5	1165.428	12.5	12.5	1	1
3	0.601815074	11	1709.2944	12.5	12.5	1	1
4	0.802420099	12.5	1942.38	12.5	12.5	1	1
5	1.003025124	12.3	1911.30192	12.7	12.6	1.007937	1.012
6	1.203630149	12.1	1880.22384	12.9	12.7	1.015748	1.024
7	1.404235173	12	1864.6848	13.2	12.8	1.03125	1.04
8	1.604840198	11.9	1849.14576	13.6	13	1.046154	1.064
9	1.805445223	12	1864.6848	14.1	13.2	1.068182	1.092
10	2.006050248	11.9	1849.14576	14.6	13.5	1.081481	1.124
11	2.206655272	12.1	1880.22384	15.1	13.8	1.094203	1.156
12	2.407260297	12.2	1895.76288	15.5	14.1	1.099291	1.184
13	2.607865322	12.2	1895.76288	16.7	14.5	1.151724	1.248
14	2.808470347	12.3	1911.30192	17.3	14.9	1.161074	1.288
15	3.009075371	12.4	1926.84096	18.1	15.4	1.175325	1.34
16	3.209680396	12.3	1911.30192	19.1	15.7	1.216561	1.392
17	3.410285421	12.4	1926.84096	20.3	16.1	1.26087	1.456
18	3.610890446	12.5	1942.38	21.6	16.6	1.301205	1.528
19	3.81149547	12.5	1942.38	22.8	17	1.341176	1.592
20	4.012100495	12.5	1942.38	24	17.6	1.363636	1.664
22	4.413310545	12.4	1926.84096	25.4	18.3	1.387978	1.748
24	4.814520594	12.5	1942.38	27	18.7	1.44385	1.828
26	5.215730644	12.5	1942.38	29	19.2	1.510417	1.928
28	5.616940693	12.6	1957.91904	31.5	20.4	1.544118	2.076

TABLE A.15: Variation of pressure drop with superficial gas velocity for 20:20:60 mixture with h_s= 15 cm and α= 7.47°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP(cm)**	**Δ P (in N/m^2)**	**h1(max), cm**	**h2(min), cm**	**r**	**R**
1	0.200605025	5	776.952	15	15	1	1
2	0.40121005	8	1243.1232	15	15	1	1
3	0.601815074	12	1864.6848	15	15	1	1
4	0.802420099	15.5	2408.5512	15	15	1	1
5	1.003025124	15.3	2377.47312	15.2	15.1	1.006623	1.01
6	1.203630149	15.2	2361.93408	15.4	15.2	1.013158	1.02
7	1.404235173	15.1	2346.39504	15.6	15.3	1.019608	1.03
8	1.604840198	15.2	2361.93408	15.8	15.5	1.019355	1.0433333
9	1.805445223	15.1	2346.39504	16.1	15.7	1.025478	1.06
10	2.006050248	15.1	2346.39504	16.4	16	1.025	1.08
11	2.206655272	15.2	2361.93408	16.7	16.3	1.02454	1.1
12	2.407260297	15.2	2361.93408	17.1	16.7	1.023952	1.1266667
13	2.607865322	15.3	2377.47312	17.5	17	1.029412	1.15
14	2.808470347	15.2	2361.93408	18	17.2	1.046512	1.1733333
15	3.009075371	15.4	2393.01216	18.7	17.5	1.068571	1.2066667
16	3.209680396	15.3	2377.47312	19.5	17.8	1.095506	1.2433333
17	3.410285421	15.4	2393.01216	20.3	18.2	1.115385	1.2833333
18	3.610890446	15.5	2408.5512	21.2	18.6	1.139785	1.3266667
19	3.81149547	15.6	2424.09024	22.2	19	1.168421	1.3733333
20	4.012100495	15.5	2408.5512	23.8	19.3	1.233161	1.4366667
22	4.413310545	15.5	2408.5512	25.6	19.5	1.312821	1.5033333
24	4.814520594	15.5	2408.5512	27.6	19.9	1.386935	1.5833333
26	5.215730644	15.5	2408.5512	29.8	20.3	1.46798	1.67
28	5.616940693	15.5	2408.5512	32.3	20.8	1.552885	1.77

TABLE A.16: Variation of pressure drop with superficial gas velocity for 20:20:60 mixture with h_s= 17.5 cm and α= 7.47°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP(cm)	Δ P (in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	5	776.952	17.5	17.5	1	1
2	0.40121005	8.5	1320.8184	17.5	17.5	1	1
3	0.601815074	12.5	1942.38	17.5	17.5	1	1
4	0.802420099	16	2486.2464	17.5	17.5	1	1
5	1.003025124	19	2952.4176	17.5	17.5	1	1
6	1.203630149	18.5	2874.7224	17.6	17.5	1.005714	1.002857
7	1.404235173	18	2797.0272	17.8	17.6	1.011364	1.011429
8	1.604840198	18	2797.0272	18	17.7	1.016949	1.02
9	1.805445223	18.5	2874.7224	18.2	17.7	1.028249	1.025714
10	2.006050248	19	2952.4176	18.5	17.8	1.039326	1.037143
11	2.206655272	18.5	2874.7224	18.9	18	1.05	1.054286
12	2.407260297	18.5	2874.7224	19.3	18.3	1.054645	1.074286
13	2.607865322	19	2952.4176	19.6	18.4	1.065217	1.085714
14	2.808470347	18.5	2874.7224	20	18.6	1.075269	1.102857
15	3.009075371	18.5	2874.7224	20.5	18.8	1.090426	1.122857
16	3.209680396	18.5	2874.7224	20.9	19	1.1	1.14
17	3.410285421	18.5	2874.7224	21.5	19.2	1.119792	1.162857
18	3.610890446	18.6	2890.26144	21.9	19.4	1.128866	1.18
19	3.81149547	18.7	2905.80048	22.5	19.7	1.142132	1.205714
20	4.012100495	18.8	2921.33952	24	20	1.2	1.257143
22	4.413310545	18.9	2936.87856	26	20.3	1.280788	1.322857
24	4.814520594	18.9	2936.87856	27.5	20.5	1.341463	1.371429
26	5.215730644	19	2952.4176	29	20.7	1.400966	1.42
28	5.616940693	19	2952.4176	31.7	20.9	1.516746	1.502857
30	6.018150743	19	2952.4176	33.5	21.4	1.565421	1.568571

TABLE A.17: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 10 cm and α= 4.61°

flow rate(m3/hr)	gas velocity(in m/s)	ΔP,(in cm)	ΔP(in N/m2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	2.8	435.09312	10	10	1	1
2	0.40121005	5.2	808.03008	10	10	1	1
3	0.601815074	9	1398.5136	10	10	1	1
4	0.802420099	8.4	1305.27936	10.3	10	1.03	1.015
5	1.003025124	8.4	1305.27936	10.8	10.1	1.069307	1.045
6	1.203630149	8.5	1320.8184	11.1	10.2	1.088235	1.065
7	1.404235173	8.3	1289.74032	11.5	10.3	1.116505	1.09
8	1.604840198	8.5	1320.8184	12	10.6	1.132075	1.13
9	1.805445223	8.5	1320.8184	12.7	10.7	1.186916	1.17
10	2.006050248	8.4	1305.27936	13.7	11	1.245455	1.235
11	2.206655272	8.4	1305.27936	14.5	11.4	1.27193	1.295
12	2.407260297	8.5	1320.8184	15.4	11.9	1.294118	1.365
13	2.607865322	8.1	1258.66224	16.2	12.1	1.338843	1.415
14	2.808470347	8.2	1274.20128	17	12.3	1.382114	1.465
15	3.009075371	8.1	1258.66224	17.9	12.5	1.432	1.52
16	3.209680396	8.4	1305.27936	18.8	12.7	1.480315	1.575
17	3.410285421	8.4	1305.27936	19.8	12.8	1.546875	1.63
18	3.610890446	8.3	1289.74032	20.8	12.9	1.612403	1.685
19	3.81149547	8.5	1320.8184	22	13	1.69230	1.75
20	4.012100495	8.5	1320.8184	23.3	13.1	1.77862	1.82
22	4.413310545	8.6	1336.35744	24.6	13.2	1.86363	1.89
24	4.814520594	8.5	1320.8184	25.9	13.4	1.93283	1.965
26	5.215730644	8.5	1320.8184	28	13.7	2.04379	2.085
28	5.616940693	8.6	1336.3574	30	13.9	2.15827	2.195

TABLE A.18: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 10 cm and α= 5.13°

flow rate(m^3/hr)	**gas velocity(in m/s)**	**ΔP,(in cm)**	**ΔP(in N/m^2)**	**h1(max) cm**	**h2(min), cm**	**r**	**R**
1	0.200605025	2.9	450.63216	10	10	1	1
2	0.40121005	5.4	839.10816	10	10	1	1
3	0.601815074	7.8	1212.04512	10	10	1	1
4	0.802420099	10	1553.904	10	10	1	1
5	1.003025124	9.3	1445.13072	10.5	10.2	1.029412	1.035
6	1.203630149	9.3	1445.13072	10.8	10.3	1.048544	1.055
7	1.404235173	9.5	1476.2088	11.3	10.4	1.086538	1.085
8	1.604840198	9.4	1460.66976	11.8	10.5	1.12381	1.115
9	1.805445223	9.4	1460.66976	12.5	10.6	1.179245	1.155
10	2.006050248	9.5	1476.2088	13.3	11	1.209091	1.215
11	2.206655272	9.3	1445.13072	13.8	11.4	1.210526	1.26
12	2.407260297	9.5	1476.2088	14.5	11.9	1.218487	1.32
13	2.607865322	9.5	1476.2088	15.3	12.4	1.233871	1.385
14	2.808470347	9.4	1460.66976	16.1	12.9	1.248062	1.45
15	3.009075371	9.5	1476.2088	17.2	13.6	1.264706	1.54
16	3.209680396	9.7	1507.28688	18.5	14.4	1.284722	1.645
17	3.410285421	9.7	1507.28688	19.5	15.1	1.291391	1.73
18	3.610890446	9.6	1491.74784	20.7	15.7	1.318471	1.82
19	3.81149547	9.7	1507.28688	21.9	16.2	1.351852	1.905
20	4.012100495	9.6	1491.74784	23.3	16.8	1.386905	2.005
22	4.413310545	9.6	1491.7478	24.5	17.3	1.41618	2.09
24	4.814520594	9.6	1491.7478	26	17.5	1.48571	2.175
26	5.215730644	9.6	1491.7478	27.5	17.6	1.5625	2.255
28	5.616940693	9.9	1538.3649	28.7	17.7	1.62146	2.32

TABLE A.19: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 10 cm and α= 11.2°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP,(in cm)	ΔP (in N/m^2)	h1(max), cm	h2(min), cm	r	R
1	0.200605025	3.5	543.8664	10	10	1	1
2	0.40121005	5.7	885.72528	10	10	1	1
3	0.601815074	8.2	1274.20128	10	10	1	1
4	0.802420099	9.6	1491.74784	10	10	1	1
5	1.003025124	11.1	1724.83344	10	10	1	1
6	1.203630149	13	2020.0752	10	10	1	1
7	1.404235173	12.1	1880.22384	10.7	10.5	1.019048	1.06
8	1.604840198	12.1	1880.22384	11.4	10.8	1.055556	1.11
9	1.805445223	12.3	1911.30192	11.8	11.2	1.053571	1.15
10	2.006050248	12.5	1942.38	12.4	11.4	1.087719	1.19
11	2.206655272	12.4	1926.84096	13	11.8	1.101695	1.24
12	2.407260297	12.5	1942.38	13.7	12.2	1.122951	1.295
13	2.607865322	12.4	1926.84096	14.3	12.7	1.125984	1.35
14	2.808470347	12.3	1911.30192	15.2	13.2	1.151515	1.42
15	3.009075371	12.3	1911.30192	15.8	13.9	1.136691	1.485
16	3.209680396	12.4	1926.84096	16.7	14.8	1.128378	1.575
17	3.410285421	12.2	1895.76288	17.7	15.4	1.149351	1.655
18	3.610890446	12.4	1926.84096	18.8	16.1	1.167702	1.745
19	3.81149547	12.5	1942.38	20	16.5	1.212121	1.825
20	4.012100495	12.4	1926.84096	21.3	17.1	1.245614	1.92
22	4.413310545	12.4	1926.84096	22.5	17.9	1.256983	2.02
24	4.814520594	12.5	1942.38	23.9	18.9	1.26455	2.14
26	5.215730644	12.4	1926.84096	25.4	19.1	1.329843	2.225
28	5.616940693	12.4	1926.8409	27.2	19.6	1.38775	2.34

TABLE A.20: Variation of pressure drop with superficial gas velocity for 40:30:30 mixture with h_s= 17.5 cm and α= 11.2°

flow rate(m^3/hr)	gas velocity(in m/s)	ΔP,(in cm)	Δ P (in N/m^2)	h1(max)	h2(min)	r	R
1	0.200605025	8	1243.1232	17.5	17.5	1	1
2	0.40121005	12	1864.6848	17.5	17.5	1	1
3	0.601815074	17.4	2703.79296	17.5	17.5	1	1
4	0.802420099	22.3	3465.20592	17.5	17.5	1	1
5	1.003025124	28	4350.9312	17.5	17.5	1	1
6	1.203630149	25.3	3931.37712	17.7	17.6	1.005682	1.008571
7	1.404235173	25.1	3900.29904	17.8	17.7	1.00565	1.014286
8	1.604840198	25.4	3946.91616	18	17.8	1.011236	1.022857
9	1.805445223	25.4	3946.91616	18.2	17.9	1.01676	1.031429
10	2.006050248	25.6	3977.99424	18.5	18	1.027778	1.042857
11	2.206655272	25.6	3977.99424	18.7	18.2	1.027473	1.054286
12	2.407260297	25.5	3962.4552	19	18.4	1.032609	1.068571
13	2.607865322	25.8	4009.07232	19.3	18.7	1.032086	1.085714
14	2.808470347	25.7	3993.53328	19.7	19	1.036842	1.105714
15	3.009075371	25.7	3993.53328	20.3	19.3	1.051813	1.131429
16	3.209680396	25.9	4024.61136	21	19.8	1.060606	1.165714
17	3.410285421	25.7	3993.53328	21.7	20.4	1.063725	1.202857
18	3.610890446	26.3	4086.76752	22.7	20.7	1.096618	1.24
19	3.81149547	26.1	4055.68944	24	21	1.142857	1.285714
20	4.012100495	26.1	4055.68944	25.2	21.3	1.183099	1.328571
22	4.413310545	26.3	4086.76752	26	22.1	1.176471	1.374286
24	4.814520594	26.3	4086.767	27	22.8	1.1842	1.422857
26	5.215730644	26.2	4071.2284	28.7	23.5	1.2212	1.4914
28	5.616940693	26.2	4071.2284	30.3	24.1	1.2572	1.5542

TABLE A.21: Variation of pressure drop with superficial gas velocity for 20:20:60 mixture with h_s= 10 cm and α= 11.2°

flow rate(m3/hr)	gas velocity(in m/s)	ΔP,(in cm)	ΔP(in N/m2)	h1(max)	h2(min), cm	r	R
1	0.200605025	5	776.952	10	10	1	1
2	0.40121005	8.8	1367.43552	10	10	1	1
3	0.601815074	13	2020.0752	10	10	1	1
4	0.802420099	11.9	1849.14576	10.3	10.2	1.009804	1.025
5	1.003025124	11.7	1818.06768	10.7	10.3	1.038835	1.05
6	1.203630149	11.7	1818.06768	11.2	10.4	1.076923	1.08
7	1.404235173	11.8	1833.60672	11.7	10.5	1.114286	1.11
8	1.604840198	11.6	1802.52864	12.1	10.8	1.12037	1.145
9	1.805445223	11.5	1786.9896	12.7	10.9	1.165138	1.18
10	2.006050248	11.8	1833.60672	13.4	11.2	1.196429	1.23
11	2.206655272	11.9	1849.14576	14.1	11.5	1.226087	1.28
12	2.407260297	11.9	1849.14576	15	11.8	1.271186	1.34
13	2.607865322	11.5	1786.9896	16	12.5	1.28	1.425
14	2.808470347	11.6	1802.52864	17.1	12.8	1.335938	1.495
15	3.009075371	11.7	1818.06768	18.2	13.2	1.378788	1.57
16	3.209680396	11.8	1833.60672	19.3	13.8	1.398551	1.655
17	3.410285421	11.8	1833.60672	20.3	14.5	1.4	1.74
18	3.610890446	11.9	1849.14576	21.2	15.1	1.403974	1.815
19	3.81149547	11.7	1818.06768	22.4	15.7	1.426752	1.905
20	4.012100495	11.6	1802.52864	23.5	16	1.46875	1.975
22	4.413310545	11.9	1849.1457	24.8	16.6	1.49397	2.07
24	4.814520594	12	1864.6848	26.4	17.1	1.54386	2.175
26	5.215730644	12.1	1880.2238	28.5	17.5	1.62857	2.3
28	5.616940693	12.1	1880.2238	32	17.9	1.78770	2.495

TABLE A.22: Variation of pressure drop with superficial gas velocity for 20:20:60 mixture with h_s= 17.5 cm and α= 11.2°

flow rate(m3/hr)	**gas velocity(in m/s)**	**ΔP,(in cm)**	**Δ P (in N/m2)**	**h1(max)**	**h2(min)**	**r**	**R**
1	0.200605025	9.4	1460.6697	17.5	17.5	1	1
2	0.40121005	15.3	2377.4731	17.5	17.5	1	1
3	0.601815074	22.5	3496.284	17.5	17.5	1	1
4	0.802420099	30	4661.712	17.5	17.5	1	1
5	1.003025124	27.4	4257.6969	17.7	17.6	1.005682	1.008571
6	1.203630149	27.5	4273.236	17.8	17.7	1.00565	1.014286
7	1.404235173	27.6	4288.7750	18	17.8	1.011236	1.022857
8	1.604840198	27.4	4257.6969	18.2	17.9	1.01676	1.031429
9	1.805445223	27.8	4319.8531	18.5	18	1.027778	1.042857
10	2.006050248	27.3	4242.1579	18.7	18.1	1.033149	1.051429
11	2.206655272	27.3	4242.1579	18.9	18.3	1.032787	1.062857
12	2.407260297	27.6	4288.7750	19.3	18.4	1.048913	1.077143
13	2.607865322	27.5	4273.236	19.6	18.6	1.053763	1.091429
14	2.808470347	27.6	4288.7750	20	18.8	1.06383	1.108571
15	3.009075371	27.6	4288.7750	20.5	19	1.078947	1.128571
16	3.209680396	28	4350.9312	20.9	19.2	1.088542	1.145714
17	3.410285421	27.9	4335.3921	21.5	19.4	1.108247	1.168571
18	3.610890446	27.9	4335.3921	21.9	19.7	1.111675	1.188571
19	3.81149547	28	4350.9312	22.5	20	1.125	1.214286
20	4.012100495	28.1	4366.4702	24	20.3	1.182266	1.265714
22	4.413310545	28.1	4366.4702	26	20	1.3	1.314286
24	4.814520594	28	4350.9312	27.5	20.7	1.328502	1.377143
26	5.215730644	28.1	4366.4702	29	20.9	1.38756	1.425714
28	5.616940693	28.1	4366.4702	31.7	21.4	1.481308	1.517143
30	6.018150743	28.1	4366.4702	33.5	21.9	1.52968	1.582857